



Analysis of Hybrid Hydrogen Systems

Final Report

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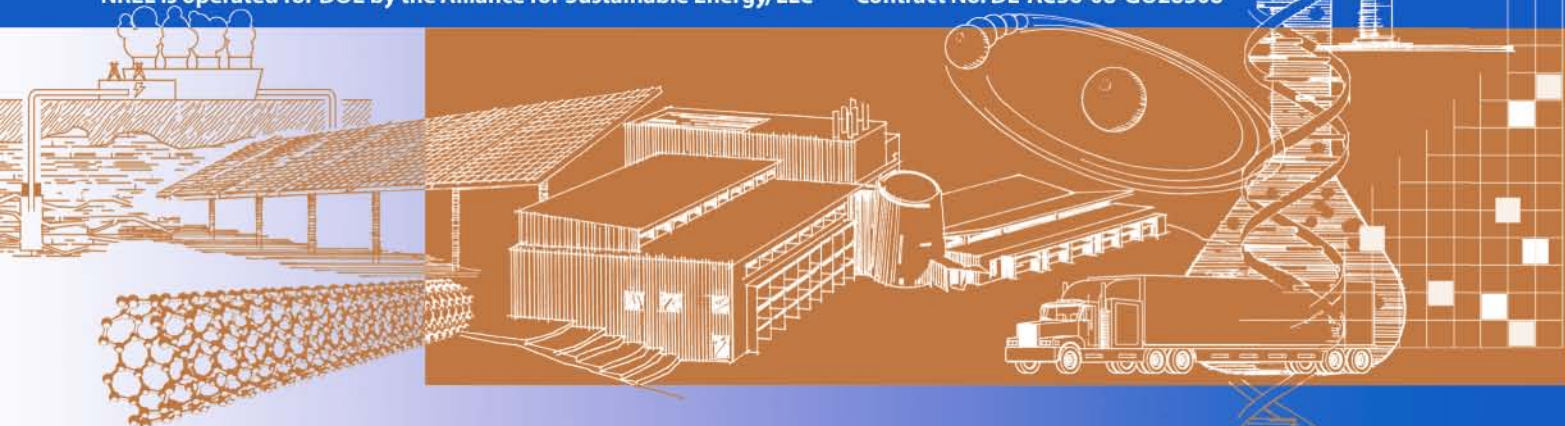
Technical Report

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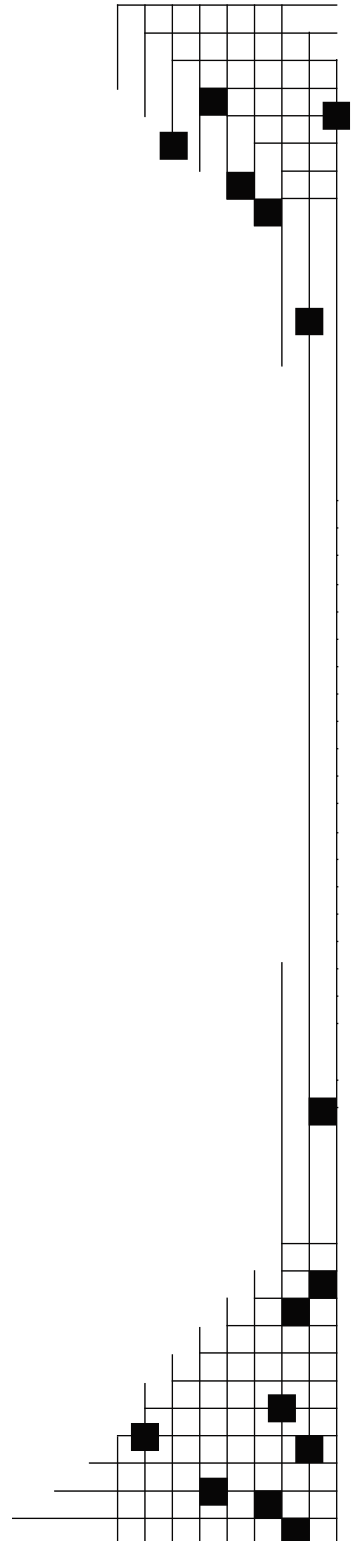
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Executive Summary

This project examined biomass pathways for hydrogen production and how they can be hybridized to support renewable electricity generation. The project considered many potential hybrid systems before narrowing the focus to two. The systems were studied in detail for process feasibility and economic performance. The best-performing system was estimated to produce hydrogen at a cost (\$1.67/kg) within range of the Department of Energy target for central biomass-derived hydrogen production, while also providing value-added energy services to the electric grid.

Of the domestic resources available for hydrogen production, biomass shows significant promise. Recent assessments have shown that more than 400 million tons of biomass currently is available annually in the United States (Milbrandt 2005). This could be converted to roughly 30 million tons of hydrogen by thermochemical processing. Thermochemical plants provide many opportunities for system integration.

The project team generated a matrix considering the combination of biomass-processing technologies and how they could be hybridized with other technologies. The matrix contained more than 100 potential binary combinations. These were ranked based on criteria such as resource availability, technology maturity, and hybridization benefits. Some of the top concepts are listed below.

- Combined wind power and biomass gasification for co-production of fuel and power

- Combined electrolysis and biomass gasification for co-production of fuel and power

- Combined coal and biomass/bio-oil gasification systems for co-production of fuel and power with carbon sequestration for both processes

- Co-location and thermal integration using steam from a nuclear reactor to feed bio-oil reforming to produce fuel

These results were further ranked using a decision matrix. Direct wind and wind-electrolyzer combinations with biomass gasification rose to the top of the decision matrix due to several factors. These selections provide renewable fuel and power, supplement grid demand, and also can take up excess electricity. The two concepts chosen for further analysis in this project can be summarized as follows.

- Direct grid leveling of intermittent wind power with an indirectly heated biomass gasification plant. The plant will produce both electricity and hydrogen.

- Using an electrolyzer in place of an air separation unit (ASU) with a directly heated fluidized-bed biomass gasifier for co-production of fuel and power.

Both of the concepts chosen for further analysis share the basic idea of combining wind-generated electricity with a biomass gasification plant. Wind availability significantly overlaps biomass resource availability, making the use of locally produced wind electricity for gasification feasible. The proposed hybrid systems attempt to do one of two things:

Fill wind energy shortfalls and feed a natural gas turbine that would be used for this peaking purpose; or

Absorb excess renewable power during low-demand hours.

The indirect gasification concepts studied could be cost competitive in the near future as value is placed on controlling carbon emissions. Carbon values of \$34 to \$40 per metric ton of carbon dioxide equivalent emission (CO₂e) make the systems studied cost competitive with steam methane reforming (SMR) to produce hydrogen. A non-hybrid biomass-to-hydrogen plant is more cost competitive in general, however, so there must be some additional value placed on peaking or sinking for these plants to make sense.

The direct gasification concept studied is unlikely to be cost competitive in the near future. High electrolyzer costs make the hybridization difficult to justify. Based on a direct replacement of the ASU with electrolyzers, hydrogen can be produced for a \$0.19 premium per kilogram. Additionally, if a nonrenewable grid mix is used, then the hybrid system is a net CO₂e emitter.

The study results and the baseline values used for comparison are summarized in the table below. A range of values is given for systems where multiple configurations and locations were studied. Cost numbers reflect current electric grid prices and demand. In the future, additional value could be placed on grid-leveling services.

Cost of Hydrogen	(\$/kg)
Steam methane reforming baseline	1.40
Indirectly heated biomass gasification to hydrogen baseline	1.64
Directly heated biomass gasification to hydrogen baseline	2.13
Electrolysis baseline	2.59
Indirect hybrid system for peaking electricity	1.81–1.88
Indirect hybrid system for sinking electricity	1.67–1.77
Direct hybrid electrolysis system	2.32

This project was reviewed by a number of biomass and power industry experts. General feedback was positive, and the team was encouraged to expand the study to include pathways such as hybrid coal gasification with sequestration. This analysis could be extended into other fuels (e.g., dimethyl ether, Fischer-Tropsch fuels), as well as to coal gasification. More-detailed process modeling to cover dynamic response and controls systems also would solidify the analysis. The systems listed above do not require significant technology breakthroughs and could be cost competitive in the near term.

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List of Acronyms

ACESA	American Clean Energy and Security Act
AEO	Annual Energy Outlook
ASU	air separation unit
BCL	Battelle Columbus
BIGCC	biomass integrated gasification combined cycle
CAES	compressed air energy storage
CNG	compressed natural gas
CO ₂ e	carbon dioxide equivalent emission
DOE	U.S. Department of Energy
EIA	Energy Information Administration
GHE	greenhouse emissions
GHG	greenhouse gas
GIS	geographic information system
GTI	Gas Technology Institute
HTS	high-temperature shift reactor
H ₂ A	Hydrogen Analysis Tool
ICC	integrated combined cycle
IGCC	integrated gasification combined cycle
ISO	independent transmission system operator
LTS	low-temperature shift reactor
MCFC	molten carbonate fuel cell
MYPP	multi-year program plan
NG	natural gas
NO _x	nitrogen oxide
NREL	National Renewable Energy Laboratory
PEM	proton exchange membrane
PO _x	partial oxidation
PSA	pressure swing adsorption
PV	photovoltaic
RTO	regional transmission organization
SC	simple cycle
SMR	steam methane reformer
SOFC	solid oxide fuel cell
SO _x	sulfur oxide
TPD	tons per day
U.S.	United States (adjective)
WGS	water-gas shift
ZnO	zinc oxide

Introduction

Hydrogen can be produced domestically from a wide variety of resources, including biomass, wind, solar, natural gas, and coal. In support of the U.S. Department of Energy's (DOE) program to develop hydrogen-production technologies for hydrogen fuel cell vehicles, this project investigates hybrid hydrogen production systems.

Of the domestic resources available for hydrogen production, biomass shows significant promise. Recent assessments show more than 400 million tons of biomass available per year in the United States (Milbrandt 2005). This could be converted to roughly 30 million tons of hydrogen by thermochemical processing. Some estimates predict that, with relatively minor changes to land management and agricultural practices, as much as 1 billion tons of biomass could be available in the future (Perlack et al. 2005). In addition to great availability, thermochemical biomass plants provide many opportunities for system integration.

The objective of this project is to evaluate the possibility of utilizing biomass' renewable and dispatchable characteristics in combination with other energy technologies to improve the efficiency, reliability, or cost of producing electricity and hydrogen from renewable energy sources. This project addresses the definition and evaluation of opportunities for combined production of hydrogen and electric power by combining biomass conversion with other hydrogen-production technologies, including wind, solar, coal, and nuclear. The National Renewable Energy Laboratory (NREL) worked with DOE to develop a list of hybridization possibilities and to perform screening of the options to identify systems that could increase the efficiency, reliability, and cost of hydrogen production, or improve the sustainability of hydrogen production from nonrenewable resources.

Analysis Methodology

Due to the open nature of the project objective, analysis was performed in consecutively refining steps. Initially, brainstorming and literature review were used to narrow the scope of research. Then researchers performed a detailed techno-economic analysis on a subset of concepts. The screening was conducted in three main stages. First, a list of possible power technologies (both renewable and nonrenewable) was generated and used for brainstorming. The results of the initial brainstorming were recorded in an idea matrix and then used to guide a literature review of existing research on hybrid systems. Many of the ideas resulting from initial brainstorming were well documented in literature, and additional ideas were added to the list of possibilities as a result of the review.

Several hybrid systems quickly rose to the top as good candidates for further investigation. These ideas were discussed extensively within the project team and then were compared using a decision matrix to narrow the options for further analysis. The most promising system concepts were analyzed in detail and examined from both a technology and economics standpoint. From a technological standpoint, hybridized system inputs and outputs were determined using Aspen Plus thermochemical software. These inputs and outputs were used by the NREL Hydrogen Analysis Tool (H2A) to determine the cost of hydrogen production for each concept.

Initial Brainstorming

During initial brainstorming the goal was to identify as many synthesis opportunities as possible between biomass conversion and other power technologies for the production of hydrogen and power. For the purposes of brainstorming, biomass conversion was assumed to include gasification, pyrolysis, direct combustion, and fermentation. Fermentation was used to describe any form of biological conversion of biomass to fuels. These five options then were compared with a list of possible power technologies, including traditional renewables such as wind and solar but also including coal, nuclear, and other technologies.

An idea matrix was constructed, with biomass technologies along the top and other power systems listed along the side. This table was used for brainstorming binary technology combinations. The summary table is shown in Figure 1 (darker cells correspond to greater perceived promise). It is based on a more-detailed numerical weighted-value matrix that can be found in Appendix A. Technologies that had large resource availability, low greenhouse gas emissions, and were relatively near term were favored in the ranking process. Initial ideation led to serious discussion of the major concepts listed below.

Combined wind power and biomass gasification for co-production of fuel and power.

Combined electrolysis and biomass gasification for co-production of fuel and power.

Combined coal and biomass gasification systems for co-production of fuel and power with carbon sequestration for both processes.

Modified, integrated, combined-cycle systems to be fed with syngas from biomass gasification for co-production of fuel and power.

Combined, direct concentrated solar and biomass gasification for fuel production.

Co-location and thermal integration using steam from a nuclear reactor to feed bio-oil reforming to produce fuel.

Co-location of biomass gasification and wastewater treatment for co-production of fuel and power.

Hybridization options	Biomass gasification	Biomass pyrolysis	Bio-oil reforming	Biomass combustion	Biomass fermentation
PV solar	0.183	0.227	0.183	0.010	0.010
Concentrated solar				0.257	
Thermal photovoltaic	0.243	0.220	0.170	0.220	0.010
Photo-electrochemical solar	0.075	0.075	0.000	0.030	0.010
Photo-bio electrolysis	0.207	0.130	0.100	0.030	0.150
Wind turbine (electrical)			0.102	0.010	0.000
Wind turbine (mechanical)	0.029	0.052	0.000	0.029	0.240
Thermal integration with nuclear		0.048		0.000	0.160
Coal/biomass co-feed		0.250	0.250		0.057
Natural gas turbine		0.257		0.257	0.071
Steam turbine		0.260			0.000
Thermal integration with coal	0.243	0.243		0.010	0.260
Alkaline fuel cell	0.000	0.000	0.000	0.000	0.000
PEM fuel cell	0.000	0.000	0.000	0.000	0.000
Phosphoric acid fuel cell	0.000	0.000	0.000	0.000	0.000
SOFC/MCFC		0.200		0.000	0.167
Hydro-electric	0.000	0.000	0.000	0.000	0.000
PEM electrolysis		0.010		0.000	0.010
Alkaline KOH electrolysis	0.237	0.000	0.227	0.000	0.000
Hybrid SOFC/SOFC electrolysis		0.160		0.030	0.010
Solid-oxide electrolysis		0.000		0.000	0.000
Thermo-chemical cycles	0.010	0.137	0.137	0.210	0.010
Wastewater treatment			0.250	0.250	

Figure 1. Idea matrix results

Literature Review Findings

A literature review was undertaken to determine the state of research with respect to biomass hybridization. The review showed that many of the concepts listed above have been or currently are being studied. There are, however, several topics where research is either sparse or nonexistent and others where significant research still is required.

All of the systems considered included some form of biomass conversion, therefore the first step was to determine the availability of biomass resources. Figure 2 shows (in green) the areas of the United States that have greater than 2,000 ton per day (TPD) of biomass available within a 50-mile radius. Multiple types of biomass were considered, including crop residues, forest residues, and primary and secondary mill residues.

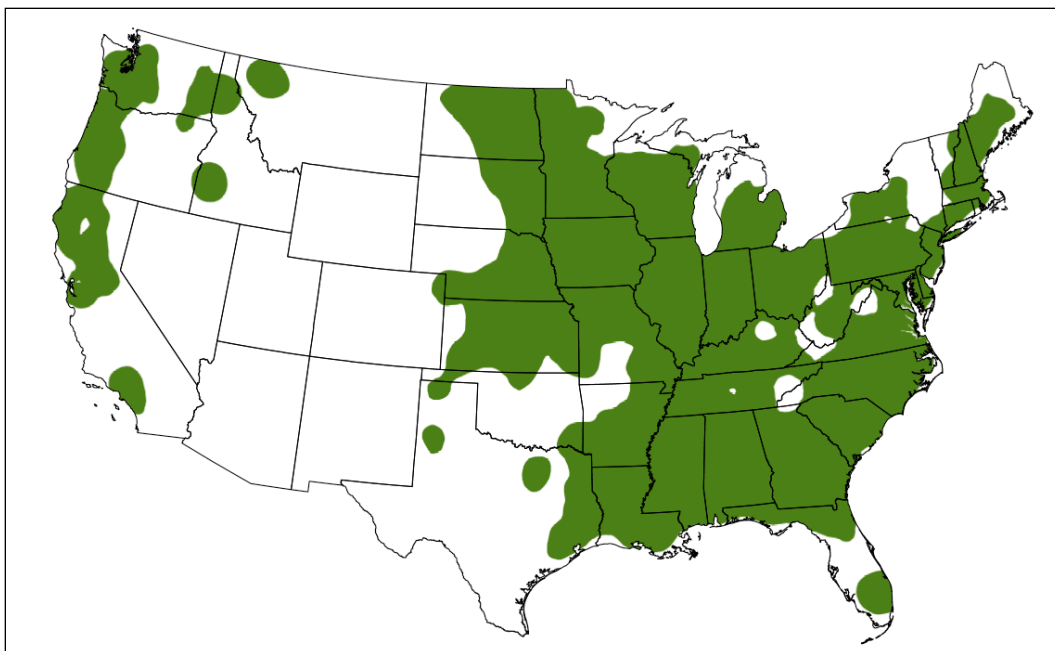


Figure 2. Biomass resources availability¹

Concentrated Solar—Biomass

Using solar energy to provide the heat for thermochemical biomass processing would reduce the environmental impact of both gasification and pyrolysis facilities. There are two main areas of research in this hybridization option—direct thermal transfer and indirect thermal transfer.



Figure 3. Parabolic (near) and tower (background) solar concentrators

In direct thermal transfer, solar concentrators are focused into the reaction chamber of a pyrolyzer or gasifier. To date, several bench-scale systems have been designed and tested with disappointing results (Lede 1999). These systems suffer from several technical problems including the amount of solar concentration needed to reach plausible reaction temperatures, solar intermittency, the need for a clear window into the reaction chamber, scalability concerns,

¹ Additional NREL biomass GIS data is available at <http://www.nrel.gov/gis/biomass.html>. Accessed November 6, 2009.

and the severe solar diffusion caused by particle movement within the reactor. In addition to technical challenges, capital costs for building a plant are expected to be significant. Although technical hurdles remain, valuable research is ongoing. A good summary of the state of technology is provided by Steinfeld in “Solar Thermochemical Production of Hydrogen—A Review” (Steinfeld 2005). Research on this topic also is ongoing at the University of Colorado.²

Indirect thermal transfer relies on heating the outer walls of a reaction chamber or heating an intermediate used for thermal storage. These systems can provide a more consistent heat source at the expense of reduced absolute temperature. Because of temperature limitations, indirect thermal transfer systems are more likely to be used for biomass pyrolysis than for gasification. One interesting option is the use of concentrated solar energy to heat molten salts, which then can be used as a pyrolyzing medium. Preliminary research suggests that, using this approach, a pyrolysis reactor could be run at steady state on solar energy alone (Adinberg, Epstein, and Karni 2004).

Whether indirect or direct thermal transfer is used, both technologies require significant solar radiation and concentration. Rough concentration ratios for trough, tower, and dish concentrators are 100, 1,000, and 3,000 suns respectively (Masters 2004). All three technologies are feasible for power generation using a heated fluid in a traditional thermal cycle. Trough concentration systems typically run at temperatures ranging from 300 to 400 °C.³ The temperatures needed for pyrolysis and gasification (500 °C and 860 °C, respectively) favor tower or dish concentration systems. Tower concentrator systems are the most likely candidate for use with a stationary chemical reactor.

The United States has significant solar resources at its disposal, but they generally are concentrated in the deserts of the southwest where biomass availability is low. Plant location is further restricted because tower concentration systems require not only high solar radiation but also large areas of flat land for construction. Transportation costs are one of the major obstacles to using biomass as an energy source, therefore the lack of local resources is problematic. Figure 4 shows the areas with biomass resources greater than 2,000 TPD within 50 miles (in green). Orange denotes solar resources in the southwestern United States of 6 kWh/m²/day or greater direct normal radiation. The solar resources are further constrained to flat sites (areas with less than 1% land slope) excluding environmentally protected lands, urban areas, and water features. In short, the orange areas show sites that might be capable of supporting a solar tower concentrator plant.

² http://aiche.confex.com/aiche/2008/preliminaryprogram/abstract_133827.htm. Accessed November 20, 2008.

³ http://www.nrel.gov/csp/troughnet/power_plant_data.html. Accessed November 8, 2009.

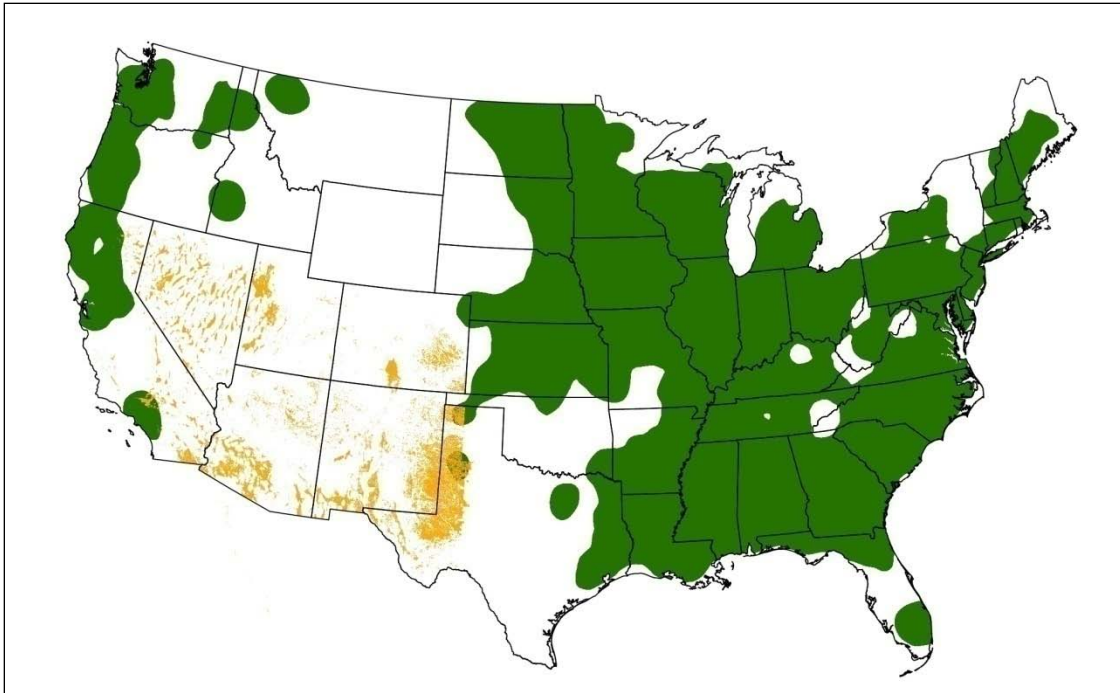


Figure 4. Solar tower plant locations versus biomass resources

Biomass gasification and pyrolysis plants typically require external electrical power for operation when the plants are optimized for fuel production. Because of this fact, many of the biomass gasification and pyrolysis research papers reviewed mentioned that the use of renewable sources of power would further add to the environmental benefits of biomass processing. Although many papers mentioned using electricity produced by renewables, no research that directly addresses the combination of intermittent photovoltaic (PV) power with thermochemical processing was found.

Use of photovoltaic electricity has several advantages over direct use of the radiation. The most significant advantage is the possibility of bringing the electricity to the biomass resources rather than having to ship biomass large distances. In addition, there is significant potential for distributed photovoltaic installation throughout the United States.⁴

The availability of low-cost solar generated electricity could be a challenge due to many factors. Peak solar radiation generally coincides with peak electricity demand, making the cost of the renewable electricity too great for cost-effective biorefinery usage. In addition, capital costs remain high for solar installations, driving up the baseline cost of solar-generated electricity.

Photo-electrochemical water splitting is a future possibility for hydrogen production via direct water splitting. This technology currently is not considered viable for hybridization because of its high cost and low efficiency, even at bench scale (Turner et al. 2008). In the future, it could provide a way to produce oxygen and hydrogen for biomass gasification without the significant electricity requirements of both cryogenic air separation units and electrolyzers.

⁴ NREL GIS data is available online at <http://www.nrel.gov/gis/solar.html>. Accessed November 8, 2009.

Integrated Gasification Combined Cycle

Biomass gasification for power generation is a more efficient route to power production than direct combustion of biomass. Extensive research has been conducted on using biomass and the syngas produced by a biomass gasifier to create power using either a gas turbine alone, a steam turbine alone, or an integrated combined-cycle approach. At least two major studies have been released by NREL directly addressing the technology, economics, and life-cycle implications of this type of hybrid power-generation technology (Craig and Mann 1996). In addition to these assessments, many biomass-to-liquid fuel studies assume that unconverted syngas is burned in a gas turbine for power generation (Larson and Celik 2005).

From a greenhouse-gas emissions standpoint, biomass-based power plants produce significantly fewer emissions than do coal or natural gas systems. Even when carbon sequestration is used on fossil-fuel plants, a biomass integrated gasification combined-cycle (IGCC) plant produces fewer atmospheric greenhouse gas emissions (Craig and Mann 1996). The major challenge for biomass-based IGCC plants is the economies of scale limitations due to biomass availability. Biomass IGCC plants typically are in the 10-MW to 60-MW range, as compared to 500-MW coal gasification plants (Electric Power Research Institute 2006).

Fuel Cell—Biomass

Coupling a fuel cell directly with the syngas output of a biomass gasification plant is a highly efficient way to produce electricity from biomass. With no moving parts and freedom from the Carnot limit, fuel cells can achieve much greater efficiencies than conventional turbines. In addition to high efficiencies, fuel cells run on a variety of fuels and typically have low maintenance requirements.

Molten carbonate (MCFC) and solid oxide (SOFC) fuel cells are the most likely candidates for combination with gasification because of their relatively low fuel-quality demands, high operating temperatures, and tolerance of carbon monoxide (Seitarides, Athanasiou, and Zabaniotou 2008). Owing to the high operating temperatures of these fuel cells (600°C to 1,000°C), it is typically most economical to use them to produce combined heat and power. Several studies have examined the possibility of combining gasification with high temperature fuel cells. Total plant electrical efficiencies of approximately 40% have been reported in literature (Wang et al. 2008).

The major obstacles to both biomass SOFC and MCFC systems are cost, syngas cleaning, and durability. Both SOFC and MCFC systems are extremely sensitive to sulfur and some of the corrosive tars produced by gasification. The cost of syngas increases with increasing purity requirements, thus cleaning the syngas for fuel cell use could be a significant burden. High-temperature fuel cells are commercially available but cost still is a major barrier to large-scale deployment, especially when combined with the high costs of gasification equipment.

Electrolysis—Biomass

Directly heated gasification systems require a source of pure oxygen if they are to be used for fuel production. Currently, plants that use oxygen produce it with cryogenic air separation units (ASU) (Ciferno and Marano 2002). Electrolysis could provide an alternative to air separation units with the added benefit of producing a pure hydrogen stream. Initial research has shown that this hybridization option could be promising from both a technical and economic perspective

(Gassner and Marechal 2008). The feasibility of producing oxygen and hydrogen with electrolysis, however, is heavily dependant on both the price of electricity and the value of the end products to the plant.

At standard temperature and pressure, an ideal electrolyzer would use 39 kWh of electricity to produce 1 kg of hydrogen. The actual state of technology limits system efficiencies to between 56% and 73% meaning that approximately 53 kWh to 70 kWh of electricity is needed for every kilogram of hydrogen produced (Kroposki et al. 2006).

Replacing a single ASU unit for oxygen production requires multiple electrolyzers. The largest commercial electrolyzer is produced by StatoilHydro (formerly NorskHydro); it has a maximum flow rate of 43.6 kg/h of hydrogen (174.4 kg/h of oxygen).⁵ A 2,000 TPD biomass gasifier would require a large bank of these electrolyzers running at full capacity. According to NREL's most recent H2A forecourt⁶ electrolysis analysis, one 174.4-kg/h electrolyzer installed with hardware costs approximately \$2.5 million. Therefore the electrolyzer bank for a 2,000 TPD gasification plant would cost significantly more than a comparable ASU unit. One possibility for addressing these high capital costs is to use enriched air for gasification rather than pure oxygen.

Along with the high capital cost of electrolysis there are other concerns with this hybridization. Water usage is a key concern with electrolyzer systems, and would be especially pronounced when combined with the generally high water requirements of biomass processes. Also, pressurized gasification plants use nitrogen from the ASU unit for pressurizing the biomass feed system. Another source of inert pressurization would be needed.

Wastewater Treatment—Biomass

Every year approximately 5.6 million dry tons of solid waste (or sludge) is produced in the United States (Bagchi et al. 2006). A significant amount of this sludge is either land filled or incinerated. Gasification could provide an alternative use for this readily available source of biomass.

The high water content of sludge is a significant challenge faced by traditional biomass gasification systems. Two options exist to overcome the water challenge. Municipal waste can be preprocessed and dried to levels acceptable to the reaction chamber before gasification, or the gasifier can be run at much lower efficiencies and the water can be vaporized in the reaction chamber itself. Using a directly heated gasifier and wet biomass would result in a significant percentage of the input carbon being burned to heat water rather than to produce syngas.

One promising alternative to traditional gasification is plasma gasification. Plasma gasification typically is done with electrically heated arc furnaces running at temperatures well above 950°C. It has been successfully used to produce high-quality syngas and power from sewage sludge in the United States, Canada, Malaysia, and Japan. The most well known of these plants is located in Japan. It produces approximately 4 MW of grid electricity by processing 138 tons of sewage sludge per day. A good overview of the state of the technology and references to existing plants can be found in "Plasma Gasification of Sewage Sludge: Process Development and Energy Optimization," by Mountouris et al. (2008).

⁵ StatoilHydro. "Hydrogen Technologies". <http://www.electrolysers.com/>. Accessed November 8, 2009.

⁶ Forecourt refers to a refueling station.

Though less glamorous, biogas digesters are another option for turning sludge into useful gas. This 100-year-old technology has been—and is—used by households in China, India, and other countries to produce natural gas for combustion in lanterns and stoves. Because the technology is simple and implementation is low cost, this approach is being adopted by many wastewater treatment facilities in the United States.

Coal—Biomass

Due to the large existing coal infrastructure in the United States, early combinations of coal and biomass involve cofiring or co-gasification of biomass with coal. Biomass can be co-fired in existing coal combustors but only in marginally small percentages due to feed problems (Wang et al. 2008). One way to address feed problems is to gasify biomass and then co-feed pulverized coal and biomass-produced syngas into existing combustors (Electric Power Research Institute 2006). Another option is to torrefy (or thermally pre-treat) the biomass, which produces a char that can be co-fed with the coal slurry. A good overview of the practical issues of dual-feed systems is provided in “Biomass Cofiring: Economics, Policy and Opportunities” (Hughes 2000). In addition to Hughes article, the DOE white paper “Biomass Cofiring: A Renewable Alternative for Utilities” (U.S. Department of Energy 2000) provides information about existing plants operating on both fuels.

Co-gasification of coal and biomass has been a focus of recent research with several good papers published in the last few years (McLendon et al. 2004; Valero and Uson 2006). These dual-feed systems help to reduce the greenhouse gas emissions of the existing coal infrastructure while maintaining economies of scale and avoiding the difficulties of finding large, reliable quantities of biomass for power generation. When significant amounts of biomass are co-fed, problems can result from increased fouling of downstream processes and high alkali content in the product ash. In the Netherlands, the Buggenum coal gasification plant has reported co-feed percentages of up to 30% with only minor changes in plant power and waste output (Electric Power Research Institute 2006). One recent paper proposed coal gasification with biomass co-feed for production of fuel and power (Cormos 2009).

Another synthesis possibility is thermal integration of biomass gasification or bio-oil reforming facilities with existing coal-fired power plants. Biomass gasification, whether directly or indirectly heated, requires a steam source that could come directly from a coal power plant. The major challenges to this type of integration are the added capital cost, the low steam temperatures relative to gasification requirements, and the mismatch in scale between biomass availability and steam production. Biomass availability could be addressed by gasification of bio-oil produced from multiple off-site pyrolysis units. Reports directly addressing the technological and economic possibilities of this system were not found during the literature review.

Nuclear—Biomass

Thermal integration and co-location of biomass processing with nuclear energy is a promising hybridization option. The presence of near carbon-neutral power and steam from the reactor could significantly increase the efficiency of a biomass plant. Charles Forsberg makes a strong case for this concept in his paper “Meeting U.S. Liquid Transport Fuel Needs with a Nuclear Hydrogen Biomass System” (Forsberg 2007).

Nuclear energy currently provides 20% of the electricity in the United States (Nuclear Energy Institute 2009). According to the Nuclear Energy Institute there are 104 nuclear reactors in the United States, and another 30 plants currently are seeking federal license approval. All of the existing plants provide a reliable source of electricity and could provide low-pressure, low-temperature steam to a biomass-processing facility. Currently, this steam is a waste stream that must be condensed after the last turbine cycle for U.S. plants. Most gasification plants would need to upgrade the steam quality before it entered the reactor, but ethanol plants could use the low-quality steam directly (Cormos 2009).

For economic reasons, to benefit from hybridization the nuclear plant would need to either be near large biomass resources or have access to low-cost barge transportation. Figure 5 shows an overlay of existing nuclear facilities (marked by black dots) versus biomass availability.

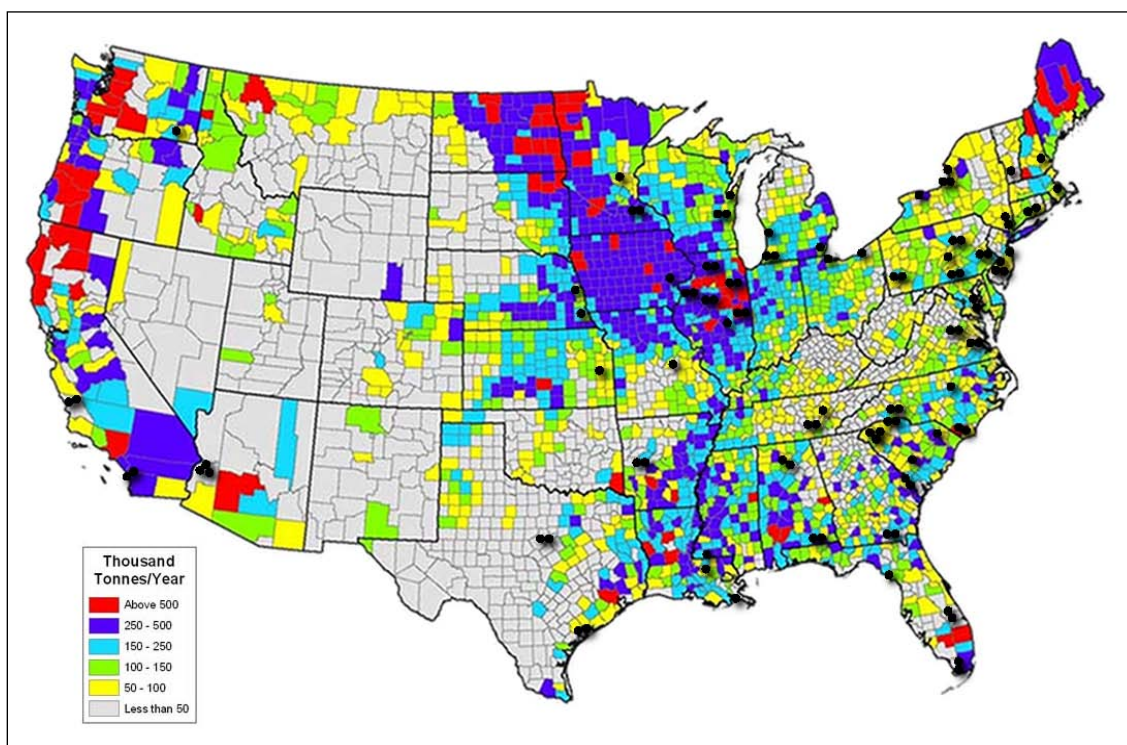


Figure 5. Nuclear plant locations versus biomass resources

Even with significant biomass resource availability, a scale mismatch between biomass availability and steam production remains. As mentioned in the coal-biomass section of this report, scale issues could be addressed by gasifying bio-oil produced from multiple off-site pyrolysis facilities (the “hub-and-spoke” concept). One way to take advantage of the scale mismatch might be to pull a slipstream of super-critical steam from the power plant steam cycle for use in gasification. No documentation or research was found on this concept. Significant plant modifications would be required for this type of integration.

Nuclear power plants rely on nonrenewable uranium resources to create heat and subsequently power. For any biomass-nuclear hybridization to be reasonable, sufficient domestic uranium resources must be available in the long term. According to Nuclear Energy Association

estimates, fuel availability is not a concern for several centuries (Price and Blaise 2002). Reserves could last significantly longer with improvements in mining technology and reactor design, and increased fuel-rod recycling.

Major challenges to such a system include the resistance of many people in the United States to the building of new nuclear plants, and U.S. security concerns involved with additional on-site processing. It might be possible to address both issues with biomass. Placing the biomass plant outside of the secure perimeter and piping the steam over the fence could negate security concerns. Creating an additional source of farm income in rural areas could go a long way towards overcoming local resistance to nuclear reactors.

Wind—Biomass

Biomass gasification and pyrolysis plants typically require external power for operation when the plants are optimized for fuel production. Because of this requirement, many of the biomass gasification and pyrolysis literature reviewed mentioned that the use of renewable sources of power would further add to the environmental benefits of thermochemical biomass processing. Although many papers mentioned using electricity produced by renewables, few examined how to directly couple intermittent wind power with thermochemical processing.

Wind turbines have quickly become a widely accepted, commercial source of renewable energy in the United States. Over the last 29 years, U.S. utilities have vastly improved their knowledge and ability to manage intermittent electricity sources. Significant issues remain, however, if large-scale wind power is pursued in the United States. These issues are addressed in detail in the Department of Energy report “20% Wind Energy by 2030” (U.S. DOE, Energy Efficiency, and Renewable Energy Laboratory 2008).

The two most significant issues with wind power are its location and its intermittency. Figure 6 shows the nation’s wind-resource distribution. The vast majority of land-based wind resources are found in the rural areas of the middle United States. To successfully utilize these resources power must be transported long distances to demand centers. Additionally, the intermittency of wind means that installing too much capacity will create grid instability unless suitable grid-leveling options are available.

Transportation of wind-generated power can be accomplished via the electrical grid or by converting the electricity to a transportable fuel. Using the national electric grid to transport the power would require significant updates to the national infrastructure. Additional high-voltage transmission lines would be needed to connect wind resources with urban areas (U.S. DOE, Energy Efficiency, and Renewable Energy Laboratory 2008). Another option is to convert intermittent electricity into a fuel. Several studies recently have been conducted on using electrolyzers to create hydrogen from wind-generated electricity (Levene, Kroposki, and Sverdrup 2006).

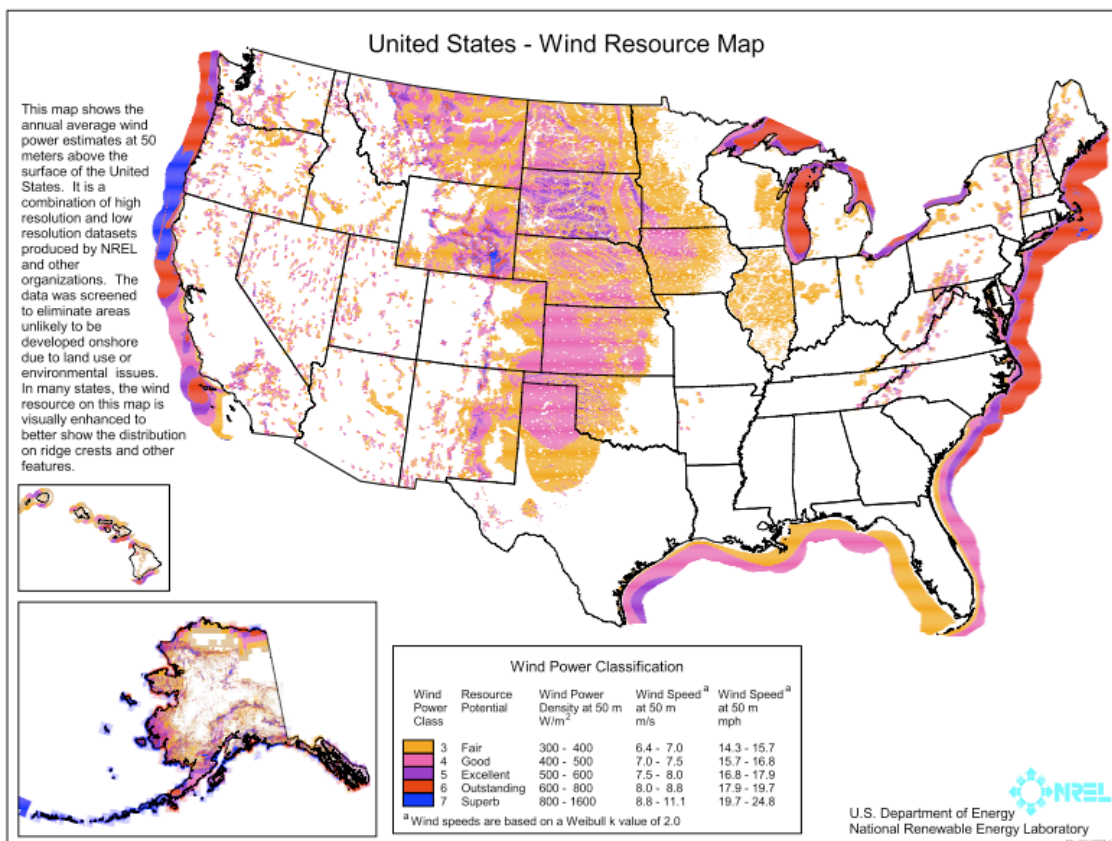


Figure 6. Wind resource availability map (U.S. DOE 2009)

Intermittency of wind electricity can cause challenges for the power grid if proper leveling options are not available. One option commonly used today is to use natural gas turbines to maintain system reliability. Turbines are readily available and can be brought online and off-line very rapidly. Another option is to use batteries and electrolyzers to store power during peak winds for use during low-wind or no-wind conditions (Fingersh 2004).

One of the only articles found to address the challenges of wind directly with biomass processing is by Paul Denholm (NREL). The proposed system would use compressed-air energy storage to store off-peak electricity generated by wind. This energy then would be used as needed by a properly designed biomass gasification plant (Denholm 2006).

It appears from Figure 6 that significant portions of national wind resources are in areas that also have biomass availability. Because of this, the shift from viewing wind electricity as an external source of electricity to trying to find direct synthesis between the two technologies appears to be a promising area of research. As a first step in this direction, we constructed maps that overlaid class 4 or better wind resources on biomass resources. For both maps the wind resources are shown in red and exclude potentially sensitive environmental lands, wind on water features, and stranded wind resources (i.e., small isolated areas). The green shading on both maps indicates that more than 2,000 TPD of the specified type of biomass is available within 50 miles. There is some “washout” of wind data because the spatial resolution on the wind resources is much finer than the biomass, so the flecks of red can be difficult to observe.

Figure 7 shows woody biomass resources versus available wind. Woody biomass includes forest residues, primary mill residues, and secondary mill residues. Primary mill residues are the bark and wood materials produced when logs are processed into lumber. Secondary mill residues consider the wood scraps from woodworking shops and factories (such as furniture manufacturers). As can be seen on the maps, there are small pockets of the northwest and northeast United States where both class 4 or greater wind and sufficient woody biomass exist for a co-located, combined system.

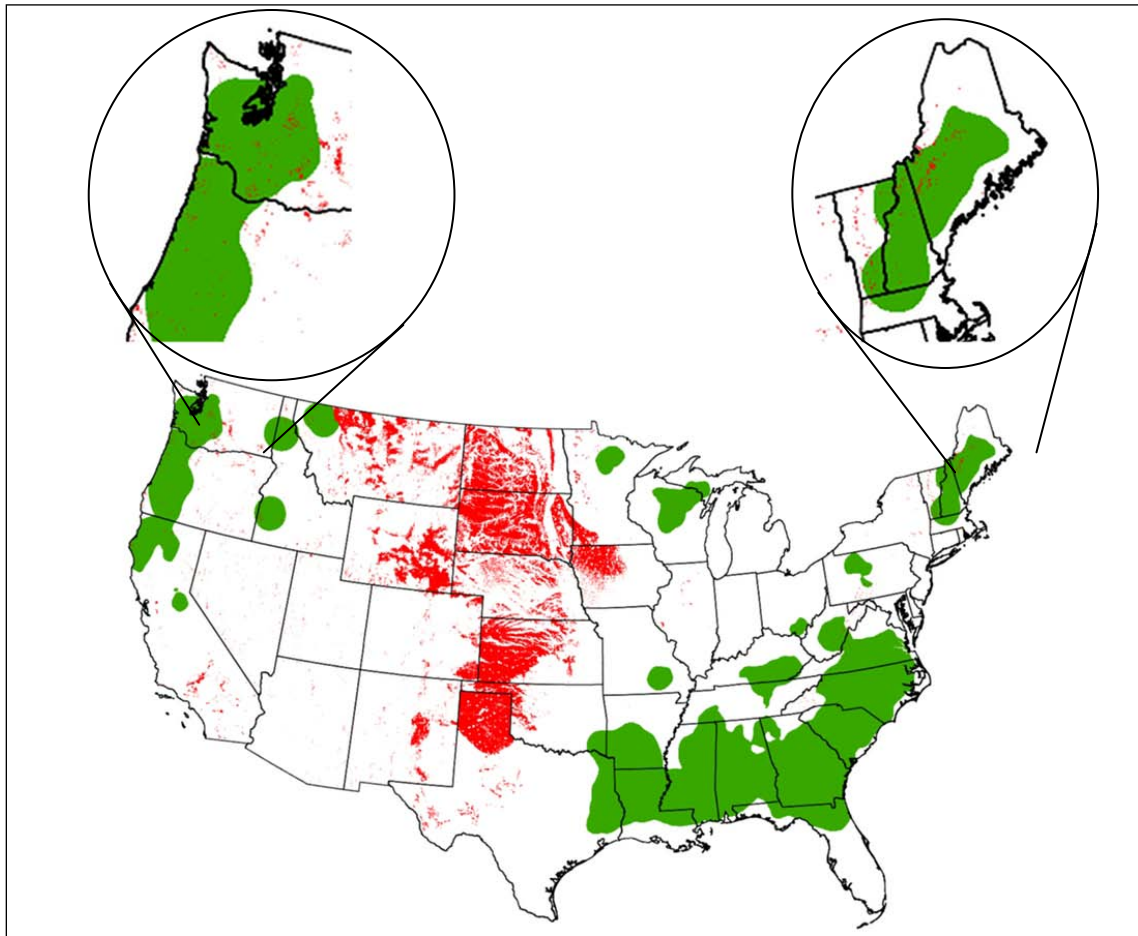


Figure 7. Wind resources versus woody biomass resources

Figure 8 shows agricultural (crop) residue biomass resources versus available wind. Crop residues considered included corn, wheat, soybeans, cotton, sorghum, barley, oats, rice, rye, canola, beans, peas, peanuts, potatoes, safflower, sunflower, sugarcane, and flaxseed. It is important to note that estimates of residue were adjusted down to allow for soil-erosion control, animal feed, bedding, and other existing farm uses (Milbrandt 2005). There is significantly more overlap of agricultural biomass with wind than woody biomass and wind.

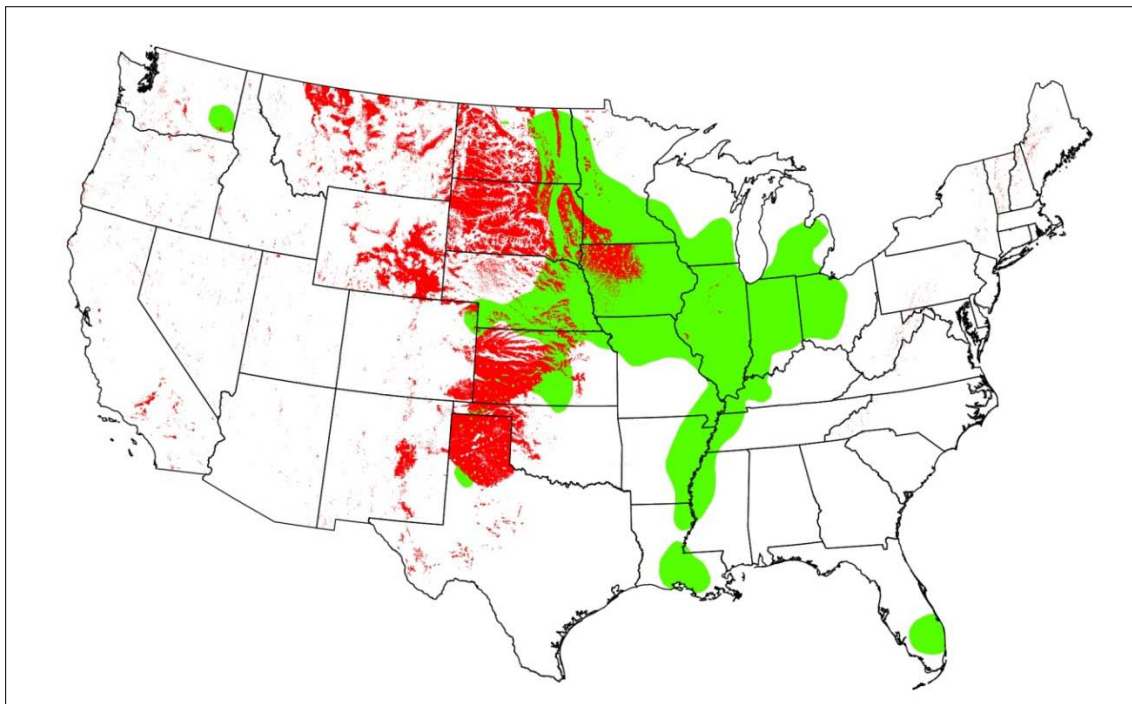


Figure 8. Wind resources versus agricultural biomass resources

Industrial Hybridization

It could be advantageous to look to industries that currently have high heat processes, biomass waste streams, or large steam requirements for synthesis possibilities. One example found in the course of literature review was the replacing of industrial gas in limekilns with synthesis gas from the gasification of hog fuel (Gribik 2007).⁷ Such research is outside of the scope of this project but might be a promising area for future research.

Concept Down Selection

Using the literature review process to refine and augment initial brainstorming, the following list of promising hybridization options was compiled.

Combined wind power and indirectly heated biomass gasification plant for co-production of fuel and power. Electric heaters could be used to increase the efficiency of hydrogen production in the plant.

Use of an electrolyzer in place of an air separation unit for a directly heated biomass gasifier for co-production of fuels and power. In addition to providing oxygen for the gasification reaction, the hydrogen produced can be used to increase fuel production of the plant. This system can be combined with wind so that additional oxygen and hydrogen can be produced during off-peak hours for use during times of high electricity demand.

Combined biomass gasification and solid oxide or molten carbonate fuel cells. The syngas produced by gasification produces power via the fuel cell. This system should provide greater overall electrical efficiency than a similar combined-cycle system.

⁷ Hog fuel is an unprocessed mix of coarse chips of bark and wood fiber.

Thermally integrated coal power plant and biomass/bio-oil gasification system. Possible opportunities for research include examining whether gasification of bio-oil could be used to improve the kinetics of coal gasification, whether waste steam could be used for gasification, and the life-cycle implications of the combination. Carbon sequestration could be used to benefit both processes.

Combined concentrated solar and molten salt biomass pyrolysis. The heat needed for biomass reforming is generated from a renewable, carbon-neutral source.

Direct, concentrated solar energy for biomass gasification.

Co-located and thermally integrated bio-oil reforming using the supercritical water from a nuclear reactor to create hydrogen or other liquid fuels (via a distributed pyrolysis system).

Co-located biomass gasification and wastewater treatment for co-production of fuel and power.

To narrow the hybrid systems (mentioned above) down to two, for further investigation, a decision matrix was constructed. A decision matrix is used to compare multiple alternatives to a chosen baseline case. For this project, all systems were compared to a biomass gasification integrated combined-cycle system. Each system then was ranked on criteria including greenhouse gas emissions, feedstock renewability and availability, reliability, cost, fuel production (syngas output per unit biomass), and the state of the technology. Rankings were input based on better (+1), worse (-1) or neutral (0). It is common practice to apply weighting factors to the criteria as needed. Because of the variety of systems involved, and to select the most generally promising systems, all weighting factors (w) for this study were equal (value of 1).

	w	Biomass ICC	Direct Wind - Gasification	Wind + Electrolyzer - Gasification	Electrolyzer - Gasification	Wastewater - Gasification	Nuclear - Gasification	Coal - Gasification	Concentrated Solar - Pyrolysis	Concentrated Solar - Gasification
Greenhouse Emissions	1	0	1	1	0	0	0	-1	1	1
Feedstock Renewability	1	0	1	1	0	0	-1	-1	1	1
Feedstock Availability	1	0	-1	-1	0	1	-1	-1	-1	-1
System Reliability	1	0	-1	-1	0	0	0	0	-1	-1
Capital Cost	1	0	-1	-1	-1	-1	1	1	-1	-1
Fuel Production Efficiency	1	0	1	1	1	0	1	0	0	0
Technology Readiness	1	0	0	0	0	0	0	0	-1	-1
New Research	1	0	1	1	0	0	0	1	0	0
Total			1	1	0	0	0	-1	-2	-2

Figure 9. Decision matrix

It can be seen from the matrix that the concentrated solar systems scored poorly due to their high cost, intermittency, and concerns with the state of the technologies involved. Coal-based systems were low on the decision matrix because of the greenhouse gas emissions involved and concerns with the scaling disparity between available local biomass resources and coal plant outputs. Electrolyzer-gasification and wastewater (or sludge) gasification are seen as alternatives to biomass combined-cycle technology. The possibility of using nuclear steam for gasification also

is an alternative to the baseline, but concerns remain about the nonrenewable fuel required for nuclear power and the scaling disparity between biomass resources and plant outputs.

Direct wind and wind-electrolyzer combinations with biomass gasification rose to the top of the decision matrix due to several factors. Little research has been done in these areas to assess economics and technical feasibility. In addition to novelty, the use of renewable wind resources to power a renewable biomass process lays the foundation for truly renewable fuel production. Finally, both cases have the potential to increase syngas and fuel yields from a given amount of biomass.

The two concepts chosen for further analysis can be stated as follows.

Direct grid leveling of intermittent wind power with an indirectly heated biomass gasification plant. The plant will produce both electricity and fuel.

Using an electrolyzer in place of an air separation unit for a directly heated biomass gasifier for co-production of fuel and power.

Both of the concepts chosen for further analysis share the basic idea of combining wind-generated electricity with a biomass gasification plant. Wind availability significantly overlaps biomass resource availability, making the use of locally produced wind electricity for gasification feasible. Additionally, gasification plants provide multiple opportunities for electricity use.

Although wind power is a promising and largely commercial renewable source of energy, its penetration of the grid poses some unique challenges. These challenges include management of intermittency with peaking units and, in the extreme case, finding use for electricity produced by wind when there is no demand. Managing intermittency will drive utilities to invest in additional peaking units and will increase the need for interruptible customers and dispatchable loads. Finally, wind in many parts of the country is a stranded resource because of a lack of grid access. Finding direct synthesis between the two technologies could allow a hybrid system to manage local intermittency or capture stranded resources.

Figure 10 shows an hourly wind profile from Norfolk, Nebraska, versus the electricity demand for the same hours as reported by the Midwest ISO. Three lines are plotted showing the wind profile at 1%, 20%, and 100% wind penetration to the market. At current levels (1%) wind energy has little if any effect on the broader energy market. At 20% wind power, significant peaking units are needed to manage the constantly changing wind generation. Dispatchable demand also could become a sought-after service to the grid at night and at any other times when high wind is not in phase with demand. Although 100% wind power is not a possibility on a regional scale, the curve is shown to emphasize the effect that wind power can have in localized pockets of the rural grid where there could be significant wind generation and low demand.

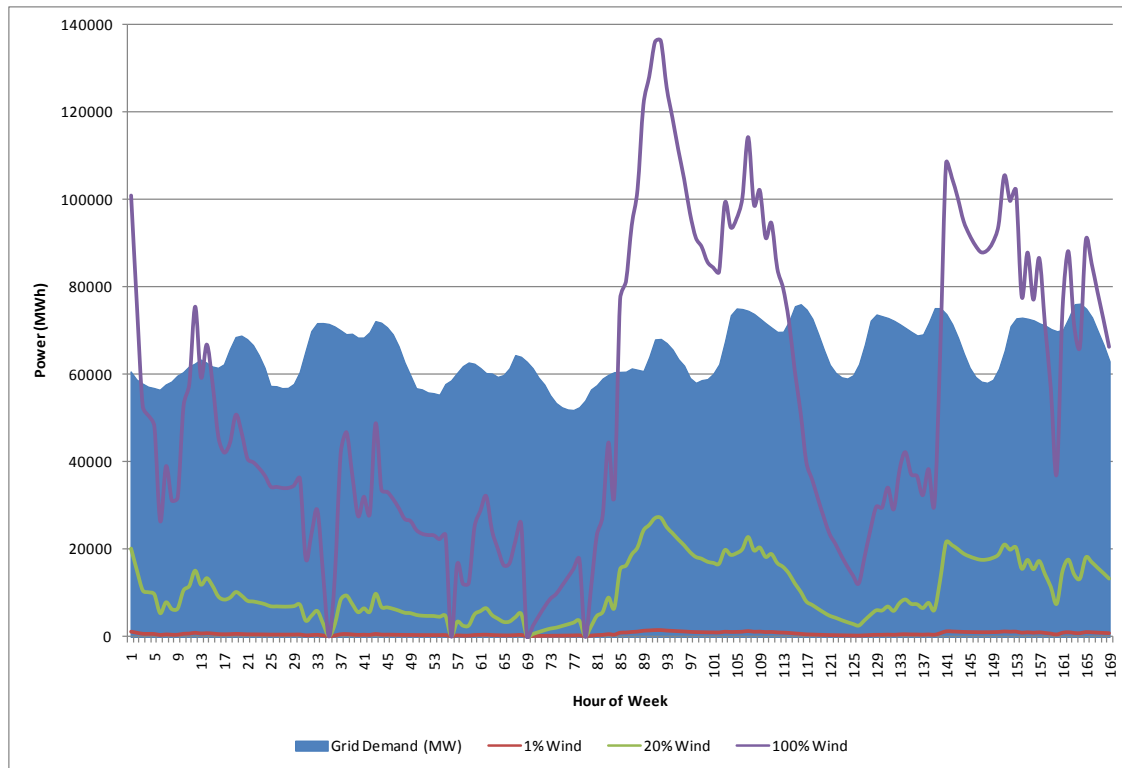


Figure 10. Wind generation versus demand for a one-week period

The proposed hybrid systems attempt to do one of two things:

Fill wind-energy shortfalls and replace the natural-gas turbines that generally are used for this peaking purpose with a renewable alternative while also generating hydrogen; or

Absorb wind-generated power when there is no demand for it. To date, most research has focused on the use of batteries, pumped hydro, hydrogen storage and fuel cells, and flywheels to store this energy for later use on the grid. Biomass gasification aided with wind electricity could “sink” this energy into fuel and provide dispatchable demand for the local utility.

Indirect Gasifier Hybrid System

This concept investigates two possible changes to a biomass-to-hydrogen plant based on indirect gasification architecture. The first modification is to allow switching between fuel production and electricity production based on grid demand. This is accomplished by routing some or all of the synthesis gas from the gasifier to a gas turbine instead of to the fuel-production reactors. In addition to power production, modifications that enable use of additional cheap or surplus electricity by the gasifier are investigated.

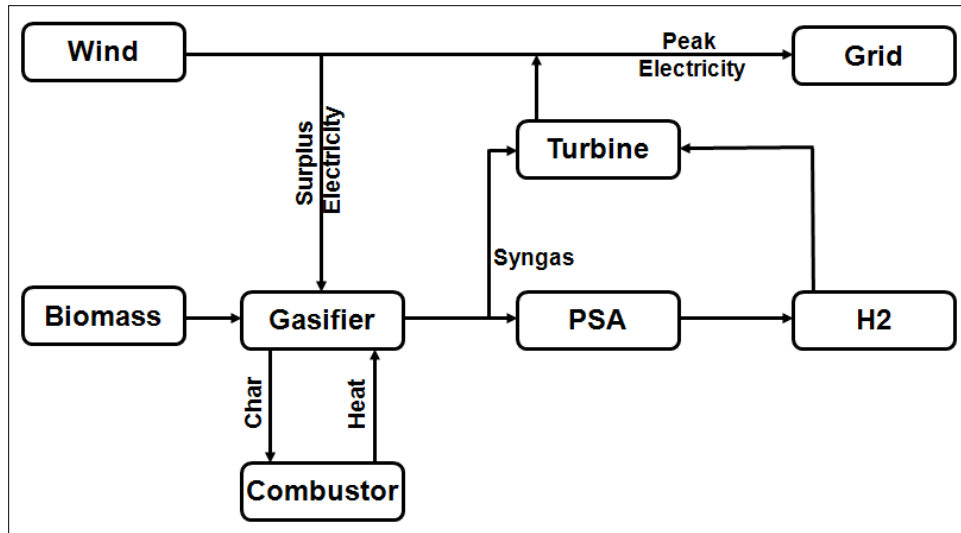


Figure 11. Indirect hybrid block diagram

Indirectly heated gasification is a two-stage fluidized-bed process where the heat needed for reaction is produced by burning char in a separate chamber to heat sand. The hot sand then is circulated through the reaction chamber to drive reaction kinetics. The layout is shown in Figure 12.

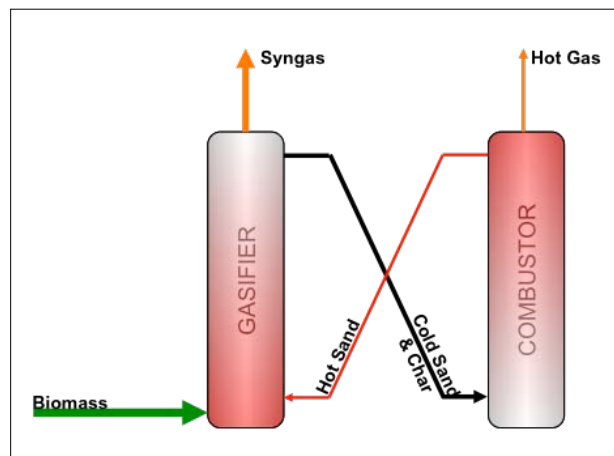


Figure 12. Indirect gasifier diagram

Electricity supplied to the gasifier during periods of low demand (lesser purchase price) will be used to heat the gasifier reaction chamber. As the temperature of the gasifier is increased, the proportions of syngas, char, and tar produced by the gasifier from a given amount of biomass change via a known relationship for a given system. Adding heat energy will create additional syngas, which will increase plant efficiency.

The ideal plant would continuously adjust both feed use and fuel production to optimize the plant economics. Electricity would be produced instead of hydrogen only when electricity was the more profitable product and vice versa. Similarly, electricity would be used for heating (or be sunk) only when electricity costs were low enough that the additional efficiency provided by the heat offsets the cost of that electricity. The feed and product selection decision is summarized in Figure 13.

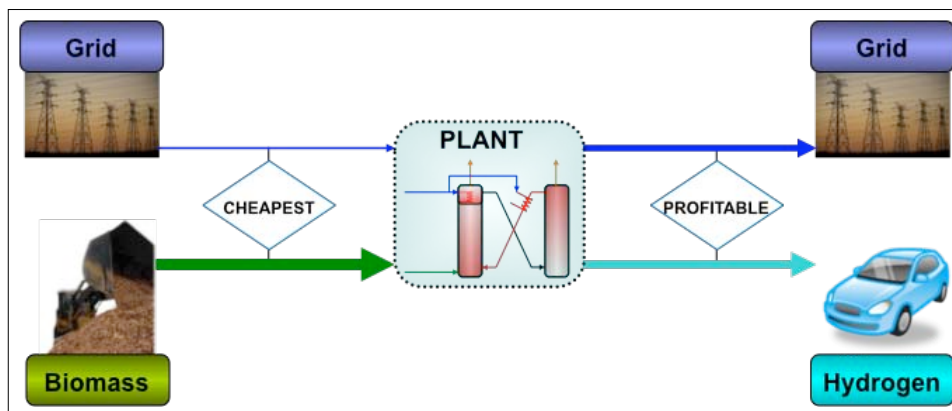


Figure 13. Concept 1 feed and product selection

Analysis of this concept was separated into the peaking and sinking modifications. The two modifications were analyzed individually to highlight the effect of each on plant economics.

Peaking Modifications

Previous NREL studies examined the possibility of using a 2,000 TPD woody biomass plant for dedicated power production and for dedicated hydrogen production (Craig and Mann 1996; Spath et al. 2005). These studies assumed steady-state operation of the biomass plant. The current analysis differs from previous studies by alternating between hydrogen production and electricity production based on market demand. This effectively combines the existing NREL Biomass-to-Hydrogen and biomass integrated gasification combined cycle (BIGCC) studies into one hybrid system. Figure 14 shows the proposed combination of the two previous studies for the current analysis.

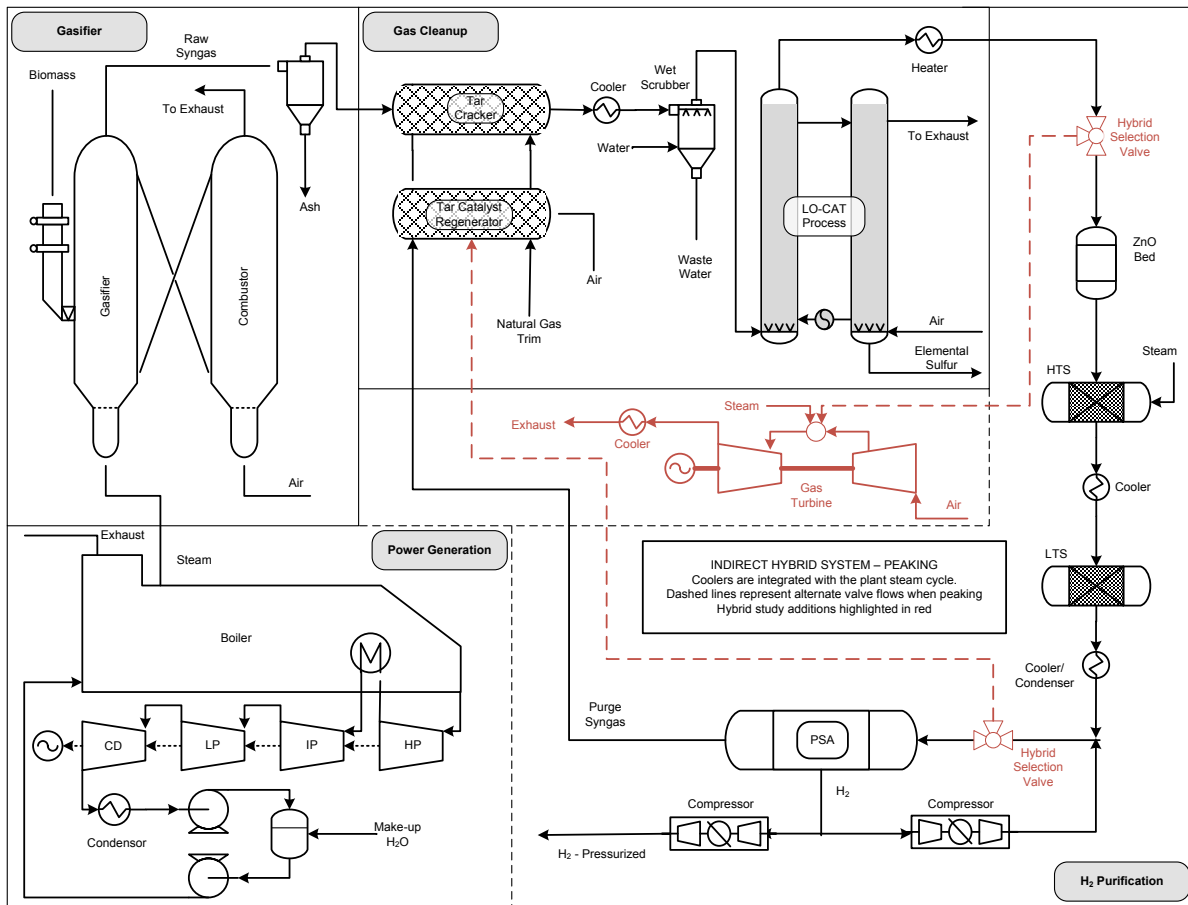


Figure 14. Proposed peaking hybrid process

The proposed plant could increase the economic promise of biomass utilization. The syngas produced by the gasifier can be used to produce hydrogen fuel or it can be used in a gas turbine to provide peaking electricity, depending on which option will maximize profit.

The synthesis gas composition available for use in all calculations is based on the biomass-to-hydrogen indirectly heated gasifier study previously completed by NREL (Spath et al. 2005). The “goal design,” flows, and specifications were used based on the suggestion of the biomass group and are referred to as the “base plant” or “base design.”

Plant Design

Multiple locations were considered for redirecting the syngas flow over to the turbine during power generation. Tar reforming is required because tars are extremely corrosive and could adversely affect the turbine combustor. Additionally, the tar represents a significant portion of the potential syngas energy, so cracking the tars converts that energy into a compatible form. This meant that the first possible location to split the syngas stream for combustion was directly after the tar reformer.

Other possible locations included:

Directly after the compression chain before the LO-CAT (hydrogen sulfide removal process) system;

Between the LO-CAT and ZnO sulfur-removal steps;

Before the pre-PSA knock-out drum; and

Directly before the PSA unit.

Table 1 shows the syngas flow, properties, composition, and approximate heating values for each option. For this analysis, syngas from the LO-CAT reactor is sent to the gas turbine. By placing the turbine after one or both of the sulfur-removal steps, emissions of SO_x can be reduced significantly.

The options of burning syngas directly prior to the PSA process or burning the hydrogen product directly in a gas turbine also were considered, but each would result in less power production due to the lower heating value of the fuel stream. Additionally, using pure hydrogen as fuel in a gas-turbine requires significant steam or nitrogen co-feeding and cooling modifications (Chiesa et al. 2005).

Table 1. Turbine Feed-Stream Possibilities

Option number	1	2	3	4	CNG
Stream number	317	326	417	420	427
Flow (kg/sec)	23.38	23.27	40.68	31.00	—
Pressure (bar)	29.3	28.3	25.5	24.8	1.0
Temp (°C)	60	48.9	43.3	43.3	15.5
–Mass Fraction (%)–					
H ₂	6.61	6.64	5.92	7.77	—
H ₂ O	0.72	0.31	24.05	0.35	—
CO	54.21	54.48	1.73	2.27	—
N ₂	0.14	0.14	0.08	0.11	1.8
CO ₂	36.04	36.22	66.96	87.87	1.3
H ₂ S	0.09	—	—	—	—
NH ₃	0.02	0.02	0.01	—	—
CH ₄	1.71	1.72	0.98	1.29	91.4
C ₂ H ₆	—	—	—	—	5.5
C ₂ H ₄	0.42	0.43	0.24	0.32	—
C ₂ H ₂	0.04	0.04	0.03	0.02	—
C ₆ H ₆	—	—	—	—	—
–Energy Content (MJ/kg)–					
LHV	14.5	14.6	7.89	10.4	48.3

The system will be binary, meaning that it produces either power or hydrogen, but not both simultaneously. Switching between the two modes presents several technical challenges due to the great degree of thermal integration in the base plant.

The most significant technical challenges identified included the following.

Downstream water-gas shift (WGS) catalyst beds are extremely sensitive to air exposure and therefore syngas flow must be maintained or the beds must be effectively sealed if shut down.

The pressure swing adsorption (PSA) off-gas from hydrogen production is burned (with a small amount of trimming natural gas) to heat the tar reformer catalyst regenerator. This energy must be replaced when the PSA is not running.

The base design includes a thermally integrated steam cycle, which is partially fed by syngas cooling steps downstream of the split location. Loss of this heat energy to the steam turbine will adversely affect the plant power system.

In the base design, pressure swing adsorption off-gas is burned with a small natural gas feed in the tar reformer catalyst regenerator. The natural gas is needed because of the low heating value of the PSA off-gas (approximately 3 MJ/kg). The energy balance and mass flows into the catalyst regenerator unit in the base design are summarized in Figure 15.

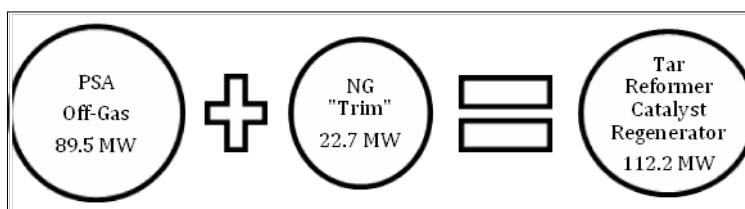


Figure 15. Tar reformer catalyst regenerator flows

If all of the syngas stream is diverted to a gas turbine for power generation for any period of time then the tar reformer will cease to function. Instead, some fraction of the syngas stream must be used to make up the lost energy flow when the PSA system is shut down. The proposed solution divides the syngas stream directly after the LO-CAT sulfur removal step so that a fraction is sent downstream through the water-gas shift reactors and then sent to the tar reformer catalyst regenerator to be burned rather than going through the PSA unit. This approach should effectively keep the sensitive WGS catalysts from air and also meet the energy demands of the tar reformer system. Additionally, it is assumed to keep all base-design systems warm and active (except for the PSA unit) until peaking is completed.

The pressure ratio between the feed and purge gas of a PSA is critical. Greater purge pressures result in significantly reduced hydrogen recovery rates. The current design uses a feed pressure of 360 psia and a purge pressure of 20 psia. Because pressure ratio is so critical to maintain proper function of the PSA unit, continuing to produce hydrogen with a lesser flow (lower pressure) does not appear to be a viable option. Because the reactor runs at low temperature and has little condensable water, a properly insulated PSA unit should be capable of being shut down for short periods.

To maintain the base-plant steam system, some portion of the exhaust gases from the gas turbine is run through a heat exchanger. The size and design of the heat exchanger will be such that it exactly replaces heat losses due to the smaller flow in the water-gas shift heat exchangers.

Modeling

Aspen Plus simulation software was used to test the feasibility of running the design scenario previously described. The existing biomass-to-hydrogen simulation was modified to include the power generation system running at steady state. Detailed discussion of the Aspen modifications can be found in Appendix B.

Splitting the syngas flow after the LO-CAT reactor was done so that the flow of trimming natural gas to the tar cracker and the flow of PSA off-gas both were replaced with syngas. Specifically, enough syngas flow was maintained through the water-gas shift reactors to exactly meet the energy demands of the tar reformer catalyst regenerator instead. The PSA is shut down during power generation. The remaining flow at the split is sent to the gas turbine. Figure 16 shows the split of syngas that meets the design requirements.

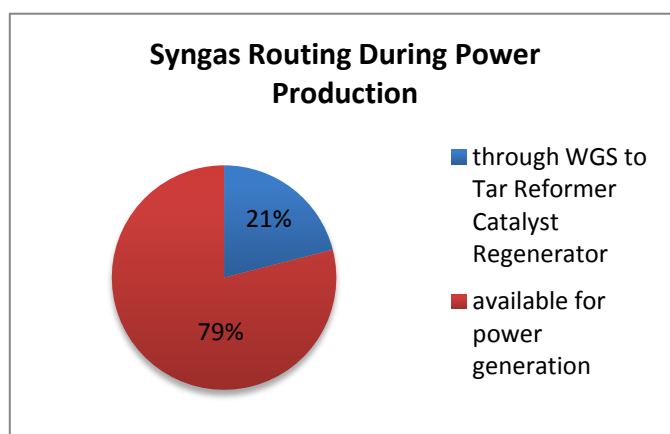


Figure 16. Syngas stream split

Given the quantity and properties of the syngas available for combustion, a gas turbine was selected for the plant. Simple cycle gas turbines typically are used for peak-shaving applications because they can be quickly brought online and off-line. Combined-cycle systems have significant start-up times due to the great amount of thermal mass involved with steam generation. One possible alternative is to run the combined-cycle system constantly but at a lower power level until more power is needed. Although the NREL BIGCC study used an integrated combined-cycle system (ICC) for power generation, a simple-cycle gas turbine was selected for this analysis. The turbines considered are summarized in Table 2.

The GE MS6001FA (6FA) simple-cycle turbine was chosen for this analysis. The 6FA and 7EA turbines have the same approximate power output, but the F class turbines are significantly more efficient than the older E class. Both the 6FA and the 7EA turbines can be run on synthesis gas. The 6B turbine initially was thought to be the proper size but was discarded as an option after the actual split fractions were determined in Aspen. The energy content in the syngas is not sufficient to use the larger 7FA family of turbines.

Table 2. GE Turbine Specifications⁸

	MS6001B	MS6001FA	MS7001EA
Output (MW)	42.1 (42)	75.9 (90)	85.1 (90)
Heat rate (kJ/kWh)	11,226	10,332	11,002
Pressure ratio	12.2:1	15.7:1	12.7:1
Mass flow (kg/s)	141	204	294
Turbine speed (rpm)	5,163	5,254	3,600
Exhaust temperature (°C)	548	603	536
Turbine inlet temperature (°C)	1,104	1,288	1,113
ICC version	S106B	106FA	107EA
ICC power output (MW)	64.3 (63)	118.1 (130)	130 (130)

Several special considerations must be made when running gas turbines on a nonstandard fuel. The turbine typically is limited by some inlet choke flow that is determined by the compressor flow and mass flow of natural gas under standard conditions. When a nonstandard (low heating value) fuel is burned, a greater fuel-feed rate typically is needed, which can cause choking of the turbine and compressor stall. The 6FA can get up to a 20% up-rate when run on syngas (from 75 MW to 90 MW) depending on the syngas composition. Allowing more mass flow through the turbine produces the extra power output. As more mass is forced through the turbine, the compression ratio increases because the turbine and compressor are coupled. According to GE technical documents (Drdar and Jones 2000) and information from a GE representative, to avoid compressor stall it is best to assume less than 14% deviation from catalogue flow rates.

In addition to high fuel-flow rates, cooling issues can arise depending on the syngas composition. Increased burner temperatures can shorten the service life of a turbine as well as drastically increasing NO_x and SO_x emissions. Based on data drawn from previous GE gasification projects (Drdar and Jones 2000) and earlier research conducted at NREL (Craig and Mann 1996), syngas compositions similar to the ones used in this study typically are humidified with steam before combustion. Steam was added to the syngas so that the final fuel gas was 20% H₂O by weight. This corresponds to a lower heating value of approximately 11 MJ/kg (174 Btu/ft³). To estimate the power output of a GE 6FA or similar gas turbine running on the available syngas composition, an Aspen Plus model for the turbine was made and inserted into the simulation. Details of the Aspen model and its calibration can be found in Appendix C.

Several important questions remain about the power generation concept that cannot be addressed using this thermodynamic model. The most important question is how the system will behave dynamically when switched between hydrogen production and power production. Answering this question requires either physical testing or dynamic simulation (with additional detailed plant design and component performance characteristics) in addition to using the model. Also, the turbine outputs for this simulation are realistic and representative, but detailed combustion analysis and testing are required to determine the plant power output more precisely.

The results of the present analysis are summarized in Table 3. When making hydrogen fuel, the plant would have the major input and output variables shown in the H₂ Production Mode column.

⁸ General Electric Company heavy-duty gas turbine products (2009).

The Power Production Mode column shows the input and output variables when the PSA unit is shut down.

Table 3. Baseline Peaking Plant Inputs and Outputs

		H₂ Production Mode	Power Production Mode
Inputs (kW)	Biomass feed	433,971	433,971
	NG feed	22,903	—
	Electricity	10,287	—
Outputs (kW)	Electricity	—	77,400
	H ₂	232,074	—

The power production value of 77.4 MW represents what a “rubber turbine”⁹ with GE F-class efficiencies and an assumed nameplate capacity of 80.6 MW could produce if all of the available syngas is utilized. A lower value of 72 MW would be produced if a GE 6FA turbine was used for the plant. This number is lower because there is a small amount of syngas available that cannot be used by a 6FA turbine.

When running in a peaking capacity, the plant has the relatively low total efficiency of 17.8%. This is compared to an efficiency of 49.7% when producing hydrogen and an expected turbine efficiency of approximately 32%. The extremely low power-production efficiency is the result of multiple factors, including the fact that 21% of the syngas stream is used to maintain the water-gas shift reactors and tar cracker rather than for power production. Additionally, a portion of the power output is used to provide power to the plant; during hydrogen production this power is provided by the grid.

The heating value of the syngas decreases after the water-gas shift reactors. This primarily is due to the conversion of carbon monoxide into other species. If, rather than sending the syngas for the tar reformer catalyst regenerator through the water-gas shift reactor before combustion, the syngas was burned immediately, then less gas would be needed (17% rather than 21% of the total syngas flow). This makes more syngas available to the rubber turbine, and 82.3 MW could be produced. This range of values (~70 MW to 85 MW) is used for sensitivity analysis in the economic model.

Alternative Design Scenario

Simple-cycle gas turbines typically are used for peaking applications because they can be cycled on and off quickly. Combined-cycle systems have significant start-up times (for example it can take more than 8 hours to cold start a 500-MW combined-cycle system; gas turbines alone can start in 12 to 15 minutes) due to the great amount of thermal mass involved with steam generation. One possible alternative is to run the combined-cycle system constantly, but at a lower power level until increased power is needed. This scenario was investigated briefly to provide data for the economic-sensitivity analysis.

⁹ The term “rubber turbine” means that the turbine size was set to exactly match the available fuel stream (as opposed to using an existing stock frame size that could not utilize all of the available fuel).

The GE 106FA combined-cycle turbine is rated at 118 MW nominal output, which is up-rated to 130 MW when the turbine is run on synthesis gas (General Electric Company 2009). Using the Aspen Plus simulations constructed and specifications for the 106FA, researchers estimated the power output possible if a combined cycle was used instead of a simple-cycle turbine.

The proposed system requires a steam turbine to operate at approximately 30% of its rated power during hydrogen production and at 100% of its rated power during power production. Although this is an extreme swing it is possible to achieve with existing technology, assuming that large efficiency losses and complex control systems are acceptable. The following approximations were made based on discussions with a GE representative.

When the steam cycle is operating at 100% rated power its thermal efficiency is 40%.

When the steam cycle is operating at 30% rated power its thermal efficiency is 30%.

Approximately 1 hour is required to ramp-up from 30% to 100% rated power.

Table 4. Peaking Plant Input/Output Combined Cycle

		H₂ Production Mode	Power Production Mode
Inputs (kW)	Biomass feed	433,971	433,971
	NG feed	22,903	—
	Plant power	17,781	—
Outputs (kW)	Power produced	—	125,710
	H ₂	232,074	—

The power production value of 125.7 MW represents what a rubber turbine with GE F-class efficiencies could produce if all of the available syngas was utilized and exhaust heat recovered. A lesser value of 110 MW is produced if a lower-efficiency steam cycle was used for the plant. When running in a peaking capacity, the plant has a significantly better total efficiency than that of the gas turbine only. The increase from 17.8% to 29% should increase the economic viability of the system; however, it still is low as compared to an efficiency of 48.9% achieved when producing hydrogen.

If the syngas was burned immediately—instead of sending the syngas for the tar reformer catalyst regenerator through the water-gas shift reactor before combustion—then less gas would be needed. This would make more syngas available to the rubber turbine and 144.98 MW could be produced. This range of values (~110 MW to 150 MW) is used for sensitivity analysis in the economic model.

Capital Costs

The additional costs to the existing Central Biomass Goal H2A analysis (Spath et. al. 2005) are summarized in Table 5. The gas-turbine numbers were taken from the *Gas Turbine World 2006 Handbook* (2006) and then adjusted to 2005 dollars using *Chemical Engineering's* Plant Cost Index.

Table 5. Capital Cost of Gas Turbines

	2006 Dollars	2005 Dollars
Uninstalled gas turbine (\$/kW)	247	231
Installation factor	1.8	1.8
Total installed cost (\$/kW)	445	417

The cost associated with additional steam turbine capacity (used in the combined-cycle alternative design scenario) was taken directly from the baseline Biomass to Hydrogen report and is shown in Table 6.

Table 6. Capital Cost of Steam Turbines

	2002 Dollars	2005 Dollars
Steam turbine (\$/kW)	474	561

Using these values and the heat exchanger prices taken from the baseline Biomass to Hydrogen report, the additional capital costs for both the simple-cycle and combined-cycle peaking systems were calculated and are summarized in Table 7.

Table 7. Concept 1 Peaking Capital Costs

Combined Cycle Additional Capital Costs		
	Simple Cycle	Combined Cycle
Turbine	\$32,249,794	\$32,249,794
Exhaust BFW preheater	\$247,729	\$247,729
Steam turbine	\$0	\$27,120,801
Total	\$32,497,523	\$59,618,324

Sinking Modifications

Indirectly heated gasification is a two-stage process in which the heat needed for reaction is produced by burning char in a separate chamber and sand is heated. The hot sand then is circulated through the reaction chamber to drive reaction kinetics. The goal of the system is for the gasifier to produce enough char to heat the reaction zone to an optimal temperature.

There is a direct correlation between the reactor temperature and the amount of syngas produced from a given amount of biomass. Higher reaction temperatures favor syngas production over char and tar production. Lower reaction temperatures cause increased tar formation and char. The relationship, as reported by previous correlations (Bain 1992), can be seen in Figure 17. Because of this correlation to temperature, indirectly heated gasifiers will reach an equilibrium temperature in the steady state.

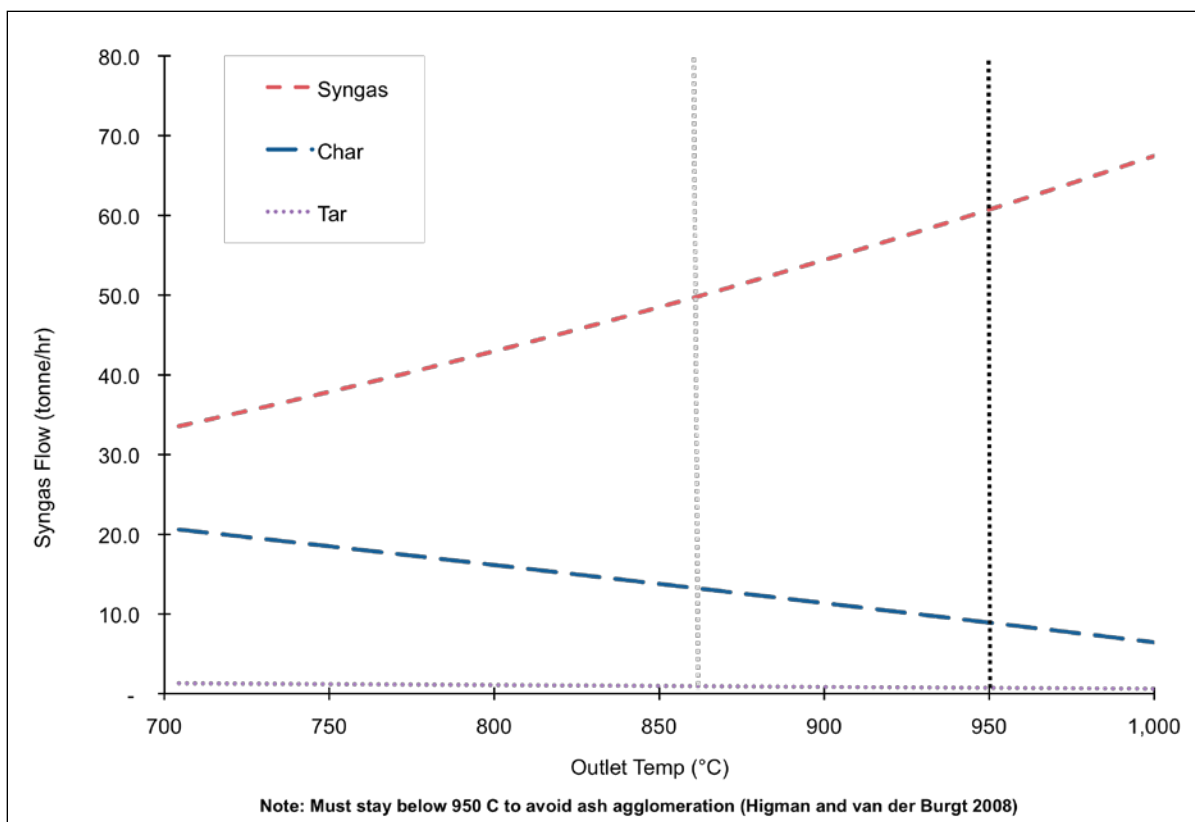


Figure 17. Indirect gasifier product flow versus temperature (2,000 TPD)

The equilibrium temperature that a gasifier reaches could be a less-than-optimal reaction temperature so that enough char can be produced to provide the required heat. Adding heat energy displaces the need for high char production and transforms some of the char and tar into additional syngas, which increases process efficiency. Alternatively, electric heating could provide a replacement for any syngas recycle or natural gas trimming currently used to push the gasifier to an optimal reaction temperature. Using excess wind power to provide electricity is one way to create a dispatchable load.

Plant Design

Initial research into how to add electric heat to the gasifier yielded interesting findings. Embedding typical resistive heaters might not be feasible in the corrosive environment of the gasifier. One possibility, however, is to use the fluidized bed itself as a “resistive element” through which the electricity passes. The Institute of Gas Technology considered this possibility in the 1970s before discarding it. The reports show that it was a technically feasible option but that electricity prices made it less cost effective than burning biomass. Using electricity selectively could change the economics.

As work began on finding pathways to electric heating of the gasifier, the NREL Biomass Center was contacted. The Biomass to Hydrogen model was being updated to include new yield correlations for the gasifier, based on data collected in the Thermochemical Process Development Unit (Kinchin and Bain 2009). The results of the updated model reveal that the gasifier does not produce enough char to maintain gasification temperatures, therefore raw syngas must be diverted and combusted to supplement the heat delivered to the gasifier by the

char combustor. A block diagram of the updated gasifier with recycled syngas is shown in Figure 18.

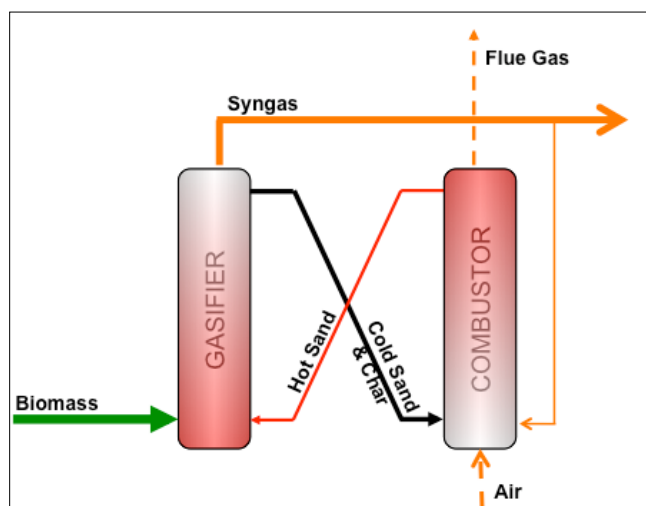


Figure 18. Updated indirect gasifier diagram

Based on the updated model, research focused on supplementing the heat delivered to the gasifier by the char combustor with electrically generated heat rather than diverting and combusting syngas. The design effort initially focused on adding heat directly to the olivine (sand) as it returned to the gasifier. This approach proved unreasonable for two reasons, listed below.

Electrically heating the olivine with currently available collar heaters (wrap-around piping, Figure 19) is not possible because commercial units are not available in the temperature and power range necessary for this application. Although a dedicated olivine heating vessel can be envisioned and assumed to exist, the efficiency and capital cost for such a unit is difficult to estimate.

The hot product gases from combusting the char and diverted syngas are used to dry the incoming biomass. If the diverted syngas is replaced with electric heaters, the olivine will receive the necessary heat to maintain gasification temperatures (approximately 870°C), but the combustion product gases used to dry the incoming biomass are limited to product gases from the char combustor alone, which are not sufficient to for drying the biomass.

If the model is to be based on currently available technology, the design most likely will employ high-power, high-temperature electric air heaters (Figure 19). If the combustion air used in the syngas and char combustors is preheated, then more heat can be delivered to the olivine per kilogram of char or syngas combusted, resulting in a reduced amount of syngas that must be recycled.

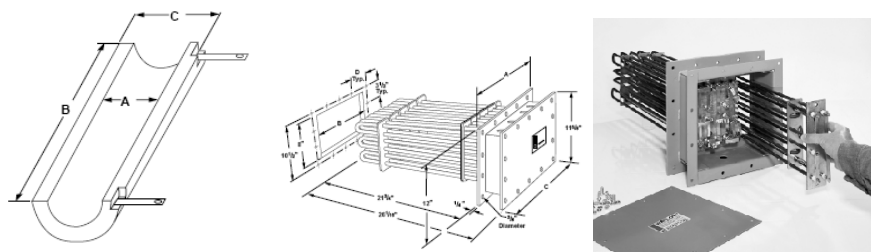


Figure 19. Collar heater, air duct heater, and air duct heater picture (Watlow 2009)

Combustion air pre-heaters actually air duct heaters that are available up to 2.2 MW. The electric air heaters also can heat air for use in the biomass dryer; therefore several design scenarios using electric heaters are possible.

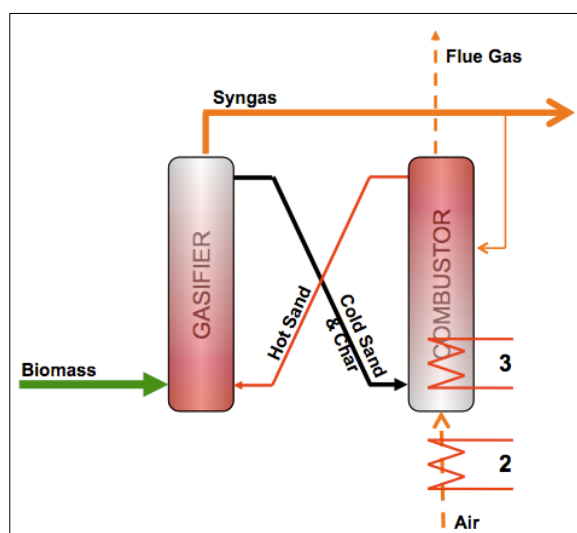


Figure 20. Final indirect gasifier sinking modifications

Basecase. No electric heat added. This design scenario is identical to the updated Biomass to Hydrogen model reported in Kinchin and Bain (2009). It provides a basis for comparing the design cases that employ electric heating.

Electric Air Heaters Only. This design case limits the electric heat assist to currently available technology (i.e., electric air duct heaters).

Electric Air Heaters and Electric Olivine Heating Vessel. This design scenario uses electric air heaters to preheat the air entering the combustor, uses electric air heaters to heat air used to dry the biomass, and includes an envisioned electric olivine heating vessel. This design scenario eliminates the need to divert any syngas for heating purposes. All heat required for gasification and biomass drying is supplied by combusting char, electrically heating air, and electrically heating olivine.

Modeling

Aspen Plus thermodynamic simulation software was used to test the feasibility of running the design scenarios previously described. As mentioned, these design scenarios were simulated using the updated Biomass to Hydrogen model (Kinchin and Bain 2009) and not the 2005 Biomass to Hydrogen model. An overview of the findings is given below.

Electric Air Heaters Only. Results of this design case are very promising. The amount of syngas that must be diverted is reduced by about 45%, and enough hot combustion products still are produced to dry the incoming biomass. Table 8 gives the detailed plant input and output values from the Aspen Plus simulation.

Table 8. Sinking Plant Input and Output with Electric Heaters Only

	Units	No Electric Heat	Electric Air Heaters Only
Biomass feed in	kg/h	83,333	83,333
Electricity in	kW	6,794	53,661
Natural gas in	kg/h	3,265	3,744
Hydrogen out	kg/h	7,134	8,005
Electric heat demand	kW	—	77,629
Electric heater power	kW	—	86,573
Power for syngas combustion air blower	kW	1,858	3,885

Electric Air Heaters and Electric Olivine Heating Vessel. As expected, results of this scenario indicate increased hydrogen yield, but at the expense of increased power requirements. Table 9 gives the detailed plant input and output values from the Aspen Plus simulation.

Table 9. Sinking Plant Input and Output with All Electric Heat

	Units	No Electric Heat	All Electric Heat
Feed in	kg/h	83,333	83,333
Electricity in	kW	6,794	110,927
Natural gas in	kg/h	3,265	4,348
Hydrogen out	kg/h	7,134	9,306
Electric heat demand	kW	0	98,820
Electric air heater power	kW	0	86,573
Electric heater tank power	kW	0	23,633
Power for syngas combustion air blower	kW	1,858	5,828

To enter these scenarios into the economic model, and so that the sinking and peaking cases could be directly compared, the Aspen results were scaled to the same baseline case that was used in the peaking analysis. Additionally, the economic model requires all inputs in terms of kilowatt, so energy content of the mass flows was calculated using the lesser heating value. The scaled and converted results are shown below for each of the two cases under consideration.

Table 10. Final Sinking Input and Output for Air Heaters Only

	Units	Baseline	Heaters
Feed in	kW	433,971	433,971
Electricity in	kW	10,287	57,154
Natural gas in	kW	22,903	27,977
Hydrogen out	kW	232,074	261,150

Table 11. Final Sinking Input and Output for All Electric

	Units	Baseline	Heaters
Feed in	kW	433,971	433,971
Electricity in	kW	10,287	114,420
Natural gas in	kW	22,903	34,367
Hydrogen out	kW	232,074	304,598

Unlike the peaking modification—which produced reduced efficiency—adding electric heat to the plant actually increases the total plant efficiency (energy in/energy out) from 49.7% to 50.3% (or 52.2% for the all-electric option). Each additional unit of energy input as electricity produces 0.56 to 0.63 units of hydrogen energy output. Therefore, while providing a dispatchable load service to the local utility, the plant actually operates more efficiently.

Capital Costs

The electric heater capital costs came from a quote provided by Watlow for its largest, high-temperature air duct heater. Based on that quote, a 2.2-MW air duct heater costs \$250,000 and has an electricity to heated air conversion efficiency of 90%. The electric air duct heaters come in 2.2-MW increments. Additional unit savings are not expected with increased size, therefore a scaling factor of 0.9 was used. Finally, an installation factor of 2.47 was assumed for all cost estimates. Based on these assumptions, the total additional capital costs for electrically heating the combustion air came to \$15.8 million.

For the all-electric heat case, the same costing assumptions were used for electric duct heaters. The cost of adding electric heating to the olivine was estimated by taking the cost of the Inconel heating coil required for delivering the given heat and multiplying that cost by 2.5. This is a very rough way to approximate the system but it provides a starting point, because little is known of the actual design of such a heater. If the system is extremely close to economical or sensitive to capital cost then sensitivity studies will be required for this value. These assumptions yield an additional capital cost of \$17.8 million. Detailed cost information for both scenarios can be found in Appendix D.

Direct Gasifier Hybrid System

The direct gasifier hybrid system concept is based on directly heated gasifier architecture. Directly heated gasifiers typically have a single combustion/reaction chamber and burn a small portion of the biomass feed to create heat. A source of pure oxygen is required for combustion if

the syngas is to be used for fuel production. Electrolysis could provide an alternative to an air separation unit, with the added benefit of producing an additional pure hydrogen stream.

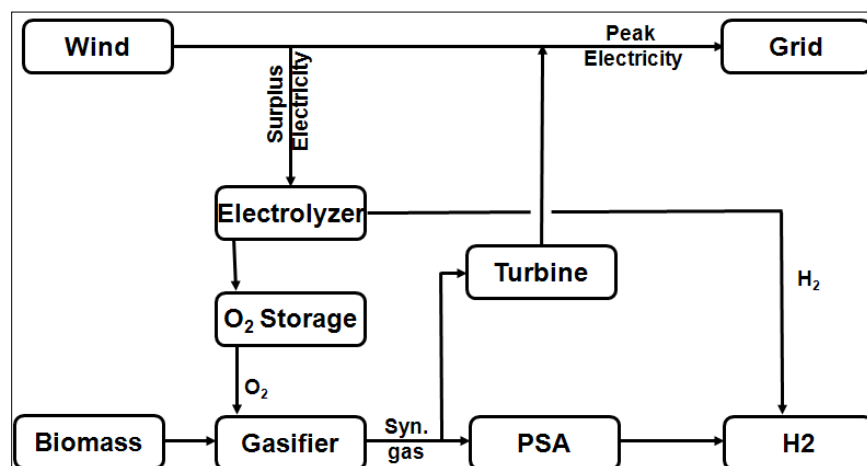


Figure 21. Direct hybrid system block diagram

To date, some research has been conducted on the feasibility of combining electrolysis with gasification (Gassner and Marechal 2008). This research concluded that the economic feasibility of this combination was greatly dependent on the price of available electricity. The proposed hybrid system directly addresses electricity price dependence by running the electrolysis system intermittently. Electricity available during periods of low demand (low purchase price) is used by electrolyzers to produce oxygen and hydrogen for use by the gasifier or stored for later use. During periods of peak electricity demand, the stored oxygen is used to create syngas rather than for running the electrolyzers.

Most directly heated gasifiers have a single combustion/reaction chamber and burn a small portion of the biomass feed to create heat. They typically are run at high pressure to improve overall plant efficiencies and reduce tar production. A source of pure oxygen is required for combustion if the syngas is to be used for fuel production so that nitrogen dilution does not affect the downstream processing. Currently, plants that use oxygen produce it with cryogenic air separation units.

The biomass-to-hydrogen plant design for a direct gasifier is similar to that of the indirectly heated gasifier previously described. The major differences between the two plants involve the addition of the air separation equipment and biomass feeding/prep equipment. An inert gas stream is needed to pressurize lock-hoppers for feeding biomass into the reactor because the gasifier is run at high pressure (approximately 24 bar). The ASUs create a stream of pure nitrogen that typically is compressed and used for this purpose. Otherwise the same cleanup processes usually can be used. Figure 22 shows an envisioned biomass-to-hydrogen pathway using the direct gasifier architecture.

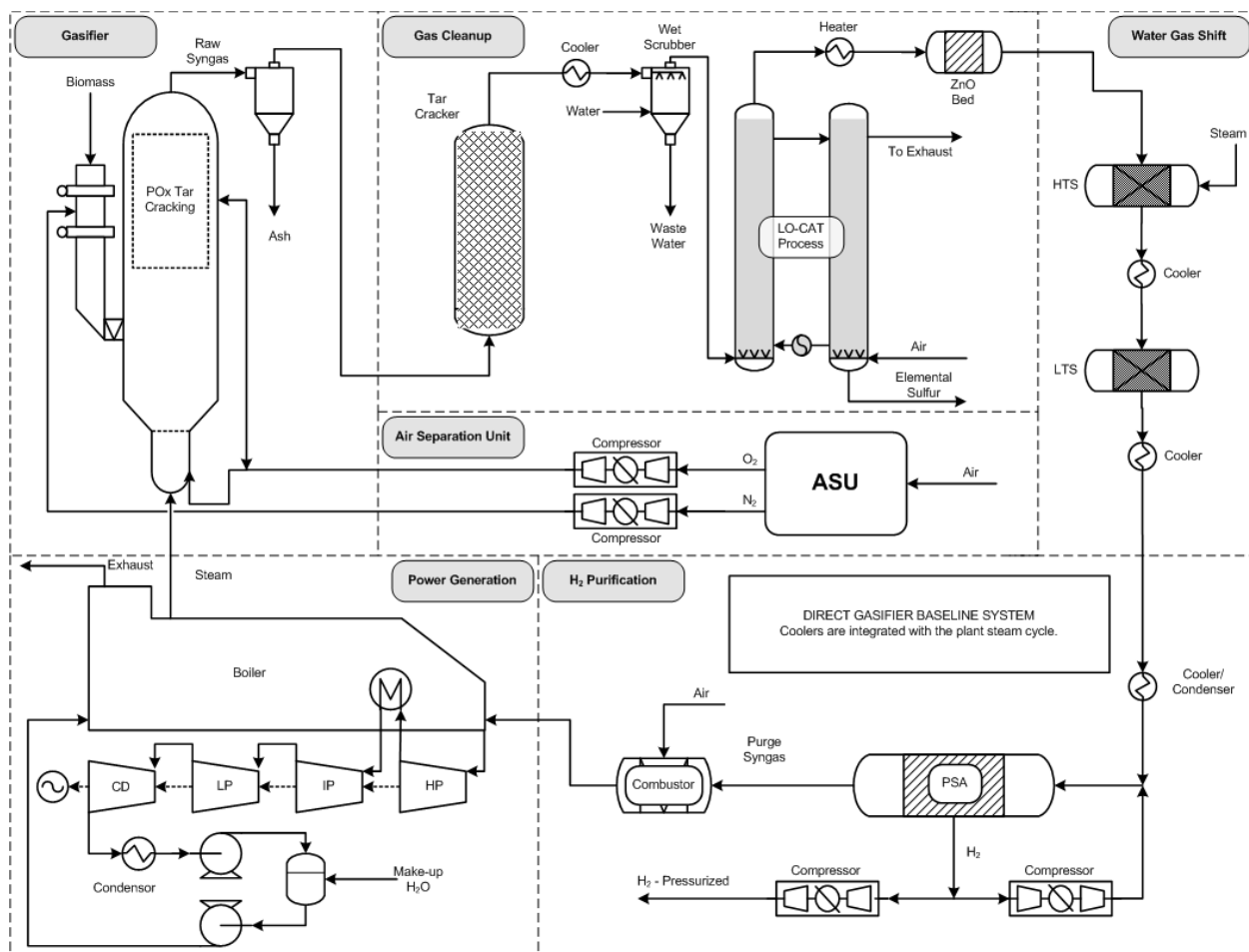


Figure 22. Direct gasifier biomass-to-hydrogen plant

Electrolysis could provide an alternative to air separation units with the added benefit of producing a pure hydrogen stream. Key changes to the plant include replacing the entire ASU with an electrolyzer bank and replacing the LO-CAT/ZnO sulfur removal steps with a two-stage Selexol plant. The sulfur removal change was driven by the need for an inert gas for feed pressurization. Selexol is a well-proven process that uses a dimethyl ether-based solvent to remove both sulfur and CO₂ from the gas stream. The envisioned plant is shown in Figure 23.

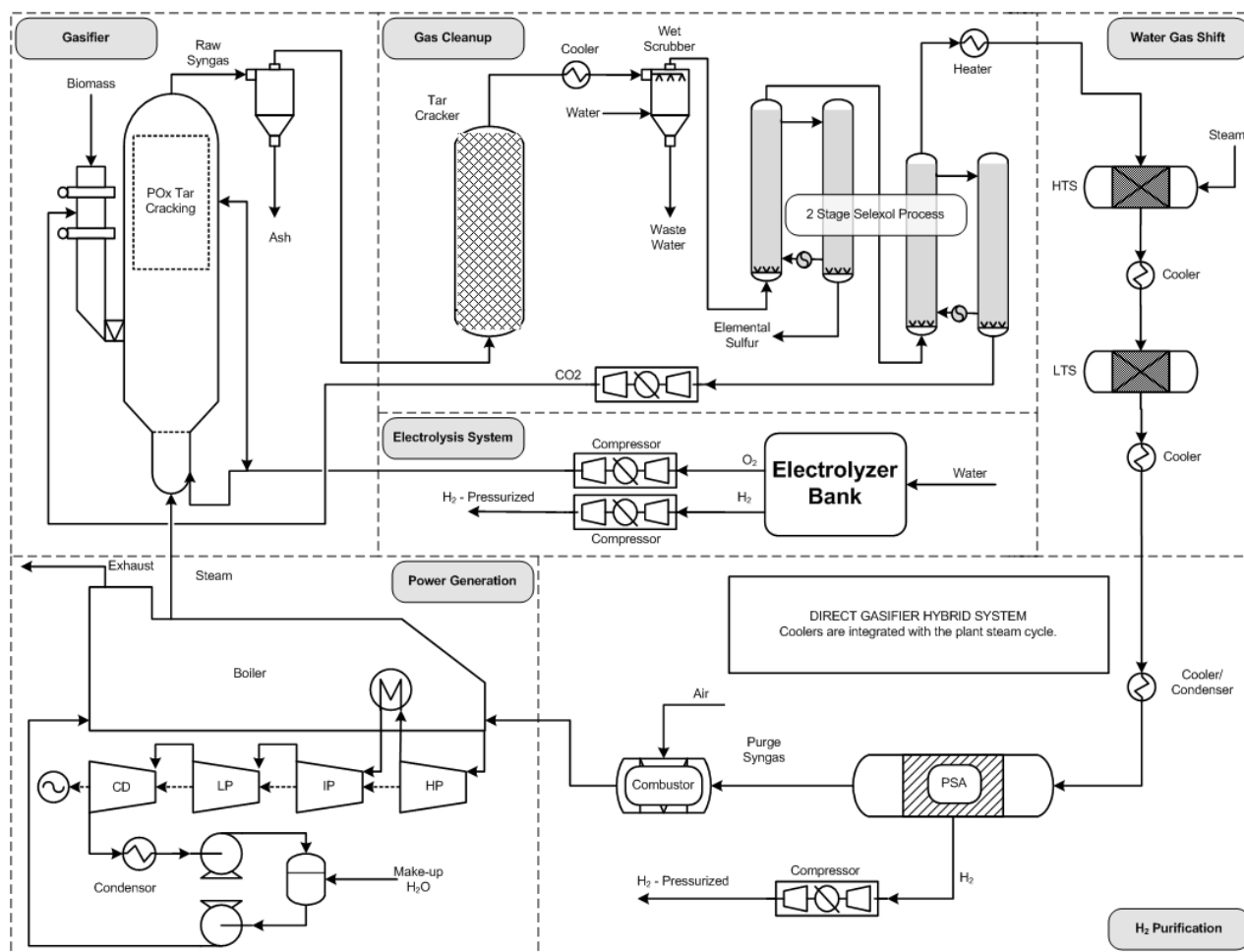


Figure 23. Direct gasifier hybrid concept

Plant Design

To replace a single ASU for oxygen production multiple electrolyzers are required. The largest commercial electrolyzer is produced by StatoilHydro (formerly NorskHydro). This electrolyzer produces a maximum flow rate of 174 kg/h of oxygen (43.6 kg/h of hydrogen). Based on published results and Aspen simulations, a 2,000 TPD fluidized-bed, biomass gasifier requires approximately 27,800 kg/hr of oxygen supplied. In this design, 160 electrolyzers running at full capacity would be needed to replace a single air separation unit. For the proposed plant to work, the Selexol process must produce enough CO_2 to replace the inert N_2 feed previously used for biomass-feed pressurization. A brief literature review found that this is in fact a possibility.

Modeling

To estimate the plant input and output values for a non-hybridized direct, fluidized-bed gasifier, an Aspen Plus model was developed based on Figure 25. The details of the model can be found in (Dean 2010). The values of biomass feed, oxygen required, water use, and electricity required obtained from this model are consistent with values published in other system simulations studies. Further refinement of the model could lead to slight changes in the plant input and output parameters, but should not be drastic enough to change the economic conclusions drawn in this study.

Given the oxygen needs of the plant, an electrolyzer bank could be designed. As noted, the electrolyzer bank will consist of 160 electrolyzers. Input and output parameters from an electrolysis plant of that size were determined from the previously released “Future Central Hydrogen Production from Grid Electrolysis” H2A analysis (U.S. Department of Energy, Hydrogen Program 2009). Based on the Aspen Plus biomass plant simulation and electrolysis H2A study, the plant input and output values for use in H2A could be successfully estimated for a combined system. The values for each individual plant and the hybrid case are given in Table 12.

Table 12. Direct Gasifier Hybrid Operating Requirements and Costs

	Units	Baseline Biomass	Baseline Electrolyzers	Hybrid System
Capacity factor	%	90	97	90
Biomass in	kg/kg H2	15.3	0	5.6
Electricity in	kWh/kg H2	0	44.7	25.4
Cooling water	gal/kg H2	170.6	293.9	174.1
Process water	gal/kg H2	5.3	2.9	3.9
Total variable operating costs	MM\$/year	\$34.8	\$117.8	\$159.9
Electricity out	kWh/kg H2	3.1	0	0
Hydrogen out	kg/day	118,344	167,360	322,440

Capital Costs

A combination of sources was used to estimate the capital investments required for both a fluidized-bed gasifier biomass-to-hydrogen case and for the proposed hybrid system. All electrolyzer costs were calculated using the future central hydrogen production from electrolysis case mentioned above. The majority of costs associated with the fluidized-bed gasifier, ASU, and feed preparation were taken from a recent publication (Jin et al. 2009). Gas cleanup costs were scaled based on previous system studies completed at NREL (Spath et al. 2005). Selexol prices were drawn from *Cost and Performance Baseline for Fossil Energy Plants* (National Energy Technology Laboratory 2007). The steam cycle and cooling costs were scaled from the 2005 Biomass to Hydrogen study based on a pinch analysis of the Aspen Plus. An overview of the costs can be found in Table 13. Detailed capital cost development information for the baseline fluidized-bed gasifier plant can be found in Appendix E and Appendix F.

Table 13. Direct Gasifier Capital Costs

Plant Area	Baseline	Hybrid
Feed preparation and handling	\$27,897,950	\$27,897,950
Gasification, tar reforming, quench	\$22,723,289	\$22,723,289
Air separation unit or electrolyzer bank	\$21,339,385	\$99,162,176
Gas cleanup	\$29,906,771	\$58,701,843
Shift and pressure swing adsorption	\$18,626,072	\$18,626,072
Steam system and power generation	\$20,423,378	\$20,423,378
Cooling water and other utilities	\$2,113,753	\$3,713,021
Buildings and structures	\$6,368,900	\$6,368,900
Total	\$149,399,497	\$257,616,627

Economic Model

The hybrid system concepts studied herein respond dynamically to fluctuations in the energy market, either absorbing or providing electricity on demand. To simulate this switching, a binary model was created for each proposed system based on a specified peaking or sinking duty. Duty is defined as the percentage of hours per year during which either sinking or peaking mode is used. The models were created in Microsoft Excel, and switch between a baseline mode and one of the hybrid modes discussed above.

Leveling of the electricity supplied to the grid from wind energy plants while producing hydrogen fuel is the goal of these concepts. However, the cost of electricity was used to determine the hybrid duty cycle instead of wind availability so that the economic promise of the concepts could be quantified.

Regional transmission organization (RTO) day-ahead prices were used for the cost-of-electricity when available. These prices represent the market value of electricity to the utilities in an area on an hourly, averaged basis. The hybrid systems must be profitable at this low price point to be able to trade electricity on the market. When day-ahead market information was unavailable, load lambda data was used in its place. Load lambda data gives the cost of production of one unit of electricity to the utility for each hour of the year.

Based on GIS research presented above (see Figure 6 and Figure 7), three areas appear to have promising quantities of both wind and biomass. These areas are the Northeast, the Midwest, and the Northwest. Cost-of-electricity data for each location is shown in Figure 24.

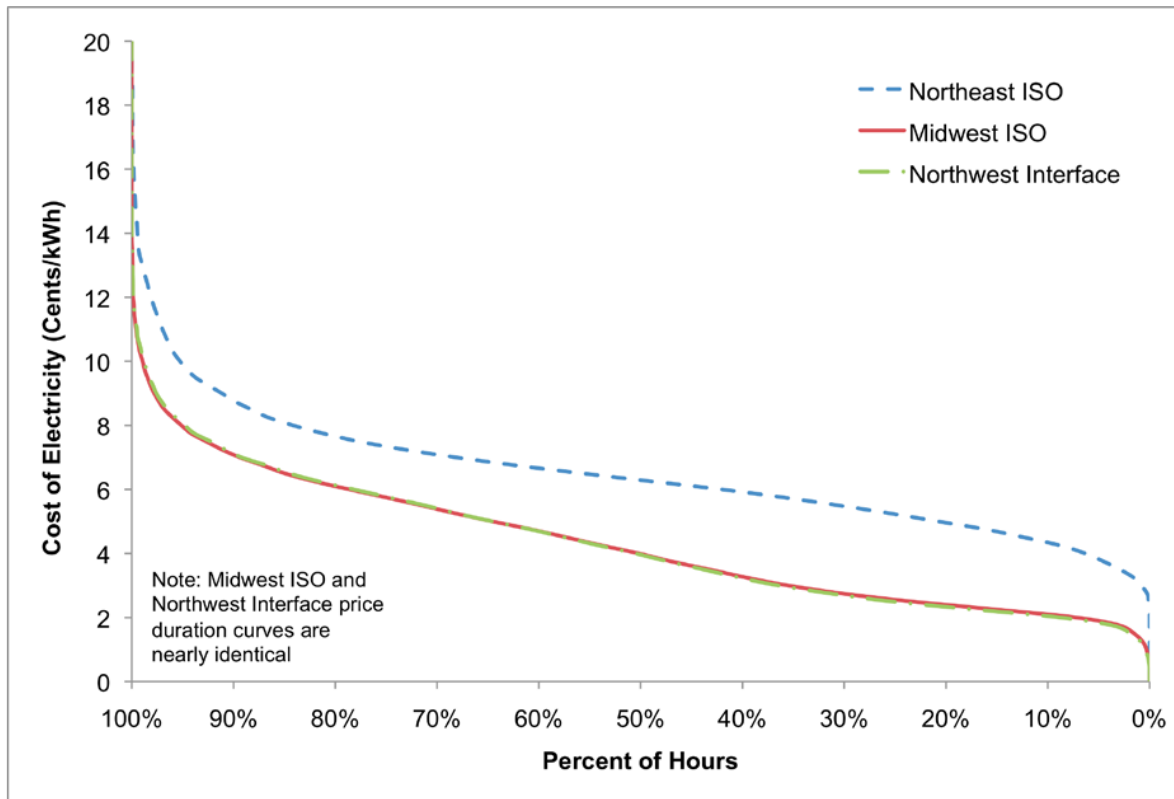


Figure 24. Regional price duration curves

Each profile shown above was based on 2007 year-end data. Costs for the Northeast were based on NE ISO day-ahead data. Midwest ISO data was directly available from the regional office itself. To estimate the Northwest, where no RTO currently exists, load lambda data was used from the Northwest Interface. More detailed information on the grid data can be found in Appendix G.

At current levels, wind penetration will not dramatically alter the cost of electricity on the regional market. In the near future wind will continue to be balanced by additional natural-gas turbines to provide the majority of peaking electricity. Based on the most recent Energy Information Administration (EIA) 2009 *Annual Energy Outlook* (AEO) report (U.S. Energy Administration 2009), electricity prices are expected stabilize at current levels and then remain unchanged for several years. Looking further ahead, the AEO predicts that the real cost of electricity will increase only \$0.014 to \$0.018 per kilowatt-hour between 2015 and 2030. This level of price increase is addressed in the sensitivity analysis found at the end of the report. In June 2009, the House of Representatives passed the American Clean Energy and Security Act of 2009 (ACESA). This is an extremely complex bill that attempts to regulate greenhouse gases with a combination of markets, efficiency programs, and incentives. Although the bill has the potential to change the rate of renewable energy deployment, it is not expected to significantly affect the AEO electricity projections (U.S. Department of Energy 2009).

Modern grid power can be roughly divided into baseload and peaking electricity. Baseload power is produced by “always-on” generation facilities like coal and nuclear plants. This currently is the lowest-cost electricity available. Peaking electricity for the most part is provided

by natural-gas turbines, which can be turned off and on quickly. Other more expensive sources of electricity such as wind and solar, however, also compete in at higher electricity prices.

For proper operation of the hybrid systems studied, the distribution of peaking or sinking needs versus baseload demand must be heterogeneous. Figure 25 shows the time distribution of costs over the course of the year for the Northeast. Because the high-priced electricity (shown in red) and low-cost electricity (shown in dark blue) occur for only a few hours at a time and appear relatively predictable, the switching proposed should be reasonable from a system-control standpoint.

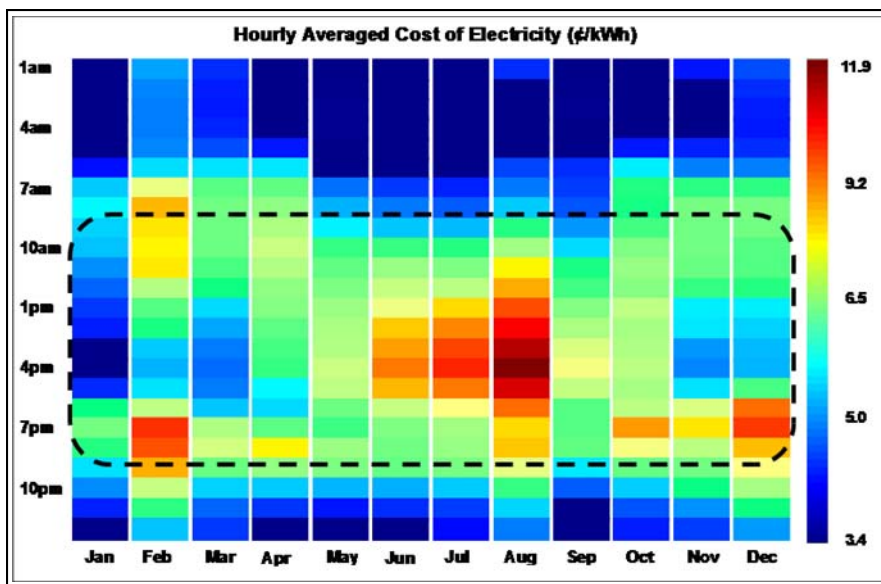


Figure 25. Time diagram of electricity cost

To assess the market potential of each of the proposed hybridizations, the yearly inputs and outputs for each plant were entered into the H2A Analysis Tool¹⁰ along with applicable capital costs. The resulting cost of hydrogen produced, in dollars per kilogram, was compared to both the Future Central Hydrogen Production via Biomass Gasification H2A results and also the Future Central Hydrogen Production from Natural Gas without CO₂ Sequestration H2A results. If the cost of hydrogen produced by a hybrid system is less than that determined for the Future Central Biomass to Hydrogen case, then the cost of additional equipment (incremental costs) for hybridization were fully offset by added income or efficiency that resulted from the hybridization.

The H2A analysis, as it was run, makes several assumptions including that current electricity prices are representative of those at the actual time of plant construction, and that sufficient market demand for hydrogen exists so that all product can be sold. The major economic assumptions are summarized in Table 14. All additional assumptions for maintenance, land, and financial variables were taken directly from the existing, published Future Biomass to Hydrogen

¹⁰ Additional information is available at http://www.hydrogen.energy.gov/h2a_analysis.html. Accessed November 8, 2009.

or Future Central Electrolysis H2A cases. Additionally, no price is associated with carbon emissions or carbon avoided and no value is added due to the dual-mode operation functionality.

Table 14. Economic Assumptions

Parameter	Value
Internal rate of return (after tax)	10%
Debt/equity	0%/100%
Plant life	40 years
Depreciation	MACRS
Depreciation recovery period	20 years
Construction period	2 years
First year	75% (25% for electrolysis)
Second year	25% (75% for electrolysis)
Start-up time	12 months
Revenues	50%
Variable costs	75%
Fixed costs	100%
Working capital	15% of total capital investment
Inflation rate	1.9%
Total taxes	38.9%
Decommissioning costs	10% of depreciable capital
Salvage value	10% of total capital investment

The last assumption, that carbon has no value, was taken into account separately as an adjustment to the H2A results. The amount of CO₂ equivalent (CO₂e) emissions per kilogram of hydrogen produced was tracked for all cases analyzed. The emissions for each hybrid system vary not only with the type of hybridization but also with the amount of time spent in each mode of operation. Values for these emissions are given with the detailed results below. A summary of the baseline values is shown in Table 15.

Table 15. CO₂e Emissions per Kilogram of Hydrogen Produced

Process	kg CO₂e/kg H₂
SMR without CO ₂ capture	11.2
Biomass to hydrogen	1.2
Hybrid systems	Varies by system

In addition to the CO₂e emissions from the plant, we assumed there was value associated with carbon emissions avoided due to the renewable nature of any fuel or electricity production. Based on the regional average grid mix, anywhere from 483 kg to 724 kg of CO₂e are emitted per kilowatt-hour of electricity produced. Table 16 shows the grid production mix and corresponding emissions for each location studied. One kilogram of hydrogen has the approximate energy equivalent of one gallon of gasoline. Because hydrogen can be used in fuel cells with much higher efficiency, however, one kilogram of hydrogen actually could offset about two gallons of gasoline. Burning two gallons of gasoline produces 17.84 kg CO₂e. These numbers were used as carbon credits for each kilogram of hydrogen or megawatt-hour of electricity produced.

Table 16. Electricity Greenhouse Emissions by Region

	Kilograms of CO ₂ e per Kilowatt-Hour ^a	NE ISO ^b	MISO ^c	NW Interface ^d
Coal	952.5	15.0%	52.2%	58.0%
Oil/petroleum	893.1	18.0%	9.3%	1.0%
Natural gas	599.2	30.0%	23.9%	19.7%
Nuclear	—	28.0%	7.9%	1.0%
Renewable	—	9.0%	6.7%	20.3%
Average kilograms of CO ₂ e per kilowatt-hour	—	483.4	723.5	679.4

a. Based on HyARC Energy Constants and Assumptions that are part of the H2A Analysis Tool.

b. Based on market reports available at <http://www.iso-ne.com/>. Accessed November 7, 2009.

c. Based on market reports available at <http://www.midwestiso.org/>. Accessed November 8, 2009.

d. Approximation based on <http://www.pacificorp.com/File/File89760.pdf>. Accessed November 8, 2009.

Results

Neither the indirect gasification–based hybrid system nor the direct gasification–based hybrid system produced hydrogen for less cost than a non-hybrid plant could. In all cases a premium was paid for hybridization that could not be offset by the increased functionality. The results for each system and the non-hybrid baseline are described below.

Indirect Hybrid System—Peaking

The proposed system switches between hydrogen production and electricity production (peaking) driven by the cost of electricity available on the grid. Based on discussions with Xcel Energy, a peaking duty of 20% was used for the analysis. Table 17 summarizes the major model inputs by region. A contract rate was used for any peaking electricity produced by the plant. This is common practice in the current electricity market and provides a premium price for dispatchable peaking assets.

Table 17. Indirect Hybrid System—Peaking H2A Inputs

	NE ISO	MISO	NW Interface
Peaking duty	20%	20%	20%
Peaking electricity value (¢/kWh)	14.0	12.0	11.5
Utility electricity cost (¢/kWh)	5.92	3.62	3.69
Utility natural gas cost (\$/nm ³)	0.32	0.32	0.32
Cost of biomass (\$/ton) ^a	48.83	48.83	48.83

a. Value taken from the *Biomass 2009 Multi-Year Research, Development and Demonstration Plan*. The 2012 target value is \$50.70 per ton of dry woody biomass in 2007 dollars. Taken to 2005 dollars with 1.09% inflation, this yields \$48.83 per ton. Available at http://www1.eere.energy.gov/biomass/pdfs/mypp_may2009.pdf. Accessed November 8, 2009.

The cost of hydrogen production in each area was calculated based on the economic inputs listed above and the plant inputs and outputs previously discussed. The results are shown in Table 18 along with the cost of hydrogen production for a baseline, non-hybridized biomass-to-hydrogen plant and a steam methane reforming (SMR) plant.

Table 18. Indirect Hybrid System—Peaking Results

	\$/kg H₂ NE ISO	\$/kg H₂ MISO	\$/kg H₂ NW Interface
SMR	1.40	1.40	1.40
Biomass-to-hydrogen baseline	1.64	1.64	1.64
Gas-turbine hybrid system	1.84	1.86	1.88
ICC hybrid system	1.81	1.85	1.87

Regardless of whether gas turbine or a swinging steam cycle is used, the additional capital costs of hybridization cannot be justified today in any of the locations studied. There is a premium of \$0.17 to \$0.24 on hydrogen produced by the proposed hybrid system compared with a non-hybrid biomass-to-hydrogen gasification plant. Areas with higher-priced electricity move this hybridization closer to economic feasibility; the best results were found in the Northeast.

The previous results assumed that carbon has no value. As carbon emissions are controlled, carbon costs could be a significant factor. If this is taken into account, then the hybrid system results substantially improve. When the hybrid plant is producing electricity the plant uses no natural gas or electricity. When the plant is producing hydrogen, however, natural gas is used for balancing the heat duty of the plant and electricity is required to run compressors. Taking the differences in carbon emissions into account, a value of \$34 to \$40 per metric ton of CO₂e makes the proposed hybrid cost competitive depending on the location and power output of the system. At \$37 per metric ton of CO₂e, the simple-cycle hybridization becomes cost competitive with a methane steam reforming plant in the Northeast. At \$34 per metric ton of CO₂e the combined-cycle system becomes cost competitive with SMR in the Northeast.

It is important to note that the baseline biomass-to-hydrogen plant requires only approximately \$23 to \$25 per metric ton of CO₂e value to be cost competitive with hydrogen produced by SMR. This means that peaking hybridization will be economically promising only when there is some value placed on the additional functionality of dual-mode operation.

A sensitivity analysis was performed for both the gas turbine and integrated combined-cycle peaking systems to characterize the effect of various technical and economic assumptions. Figure 26 and Figure 27 show the results for both the price of hydrogen and the greenhouse gas emissions. The sensitivity analysis for the gas turbine system showed that capital costs, the cost of biomass, and the price at which peaking electricity can be sold are key inputs. Capital costs—when varied +/-30%—cause the cost of hydrogen to vary by \$0.25. The cost of biomass was varied from \$40 per ton to \$60 per ton based on projections in the multi-year program plan (MYPP) (based on HyARC Energy Constants and Assumptions that are part of the H2A Analysis Tool), and causes the cost of hydrogen to vary by less than \$0.20. The price of peaking electricity and the plant capacity factor also cause fluctuations of less than \$0.20. Changes in the cost of electricity bought by the plant, the cost of natural gas, peaking power output, and turbine peaking duty cause hydrogen costs to vary less than 5%.

Greenhouse gas emissions are affected by the gas-turbine peaking duty. During hydrogen production, a natural gas trim is used to maintain the tar cracker catalyst regenerator temperature. When producing peaking electricity, the natural gas trim is replaced with synthesis gas. This switch means that higher peaking duty results in less natural gas use and therefore fewer net emissions.

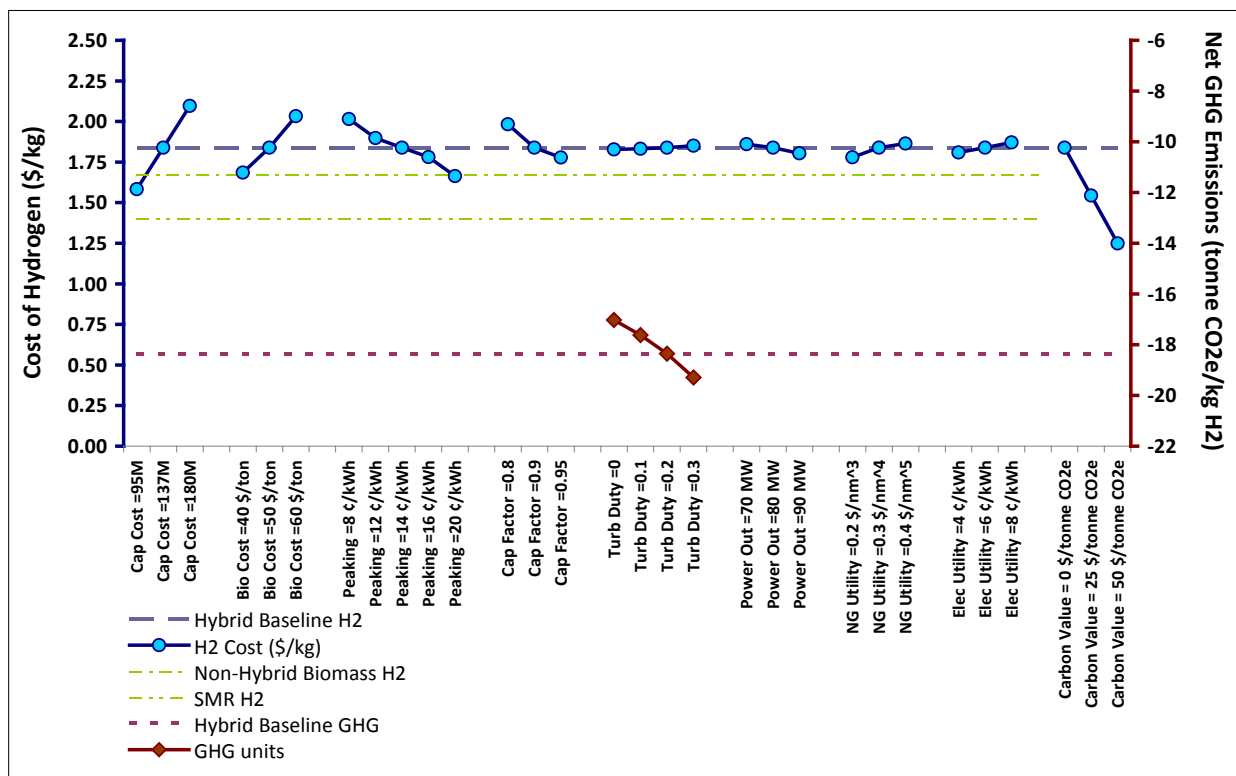


Figure 26. Gas-turbine peaking sensitivity analysis, Northeast ISO

For the integrated combined-cycle peaking system the results were significantly more volatile than for the gas-turbine system. Most of the trends discussed for the gas-turbine system held but were amplified. One difference was that the peaking duty had a much more significant effect on the cost of hydrogen, resulting in variations of up to \$0.22. This shows that the cost of hydrogen produced is extremely sensitive to the amount of peaking power produced by the system.

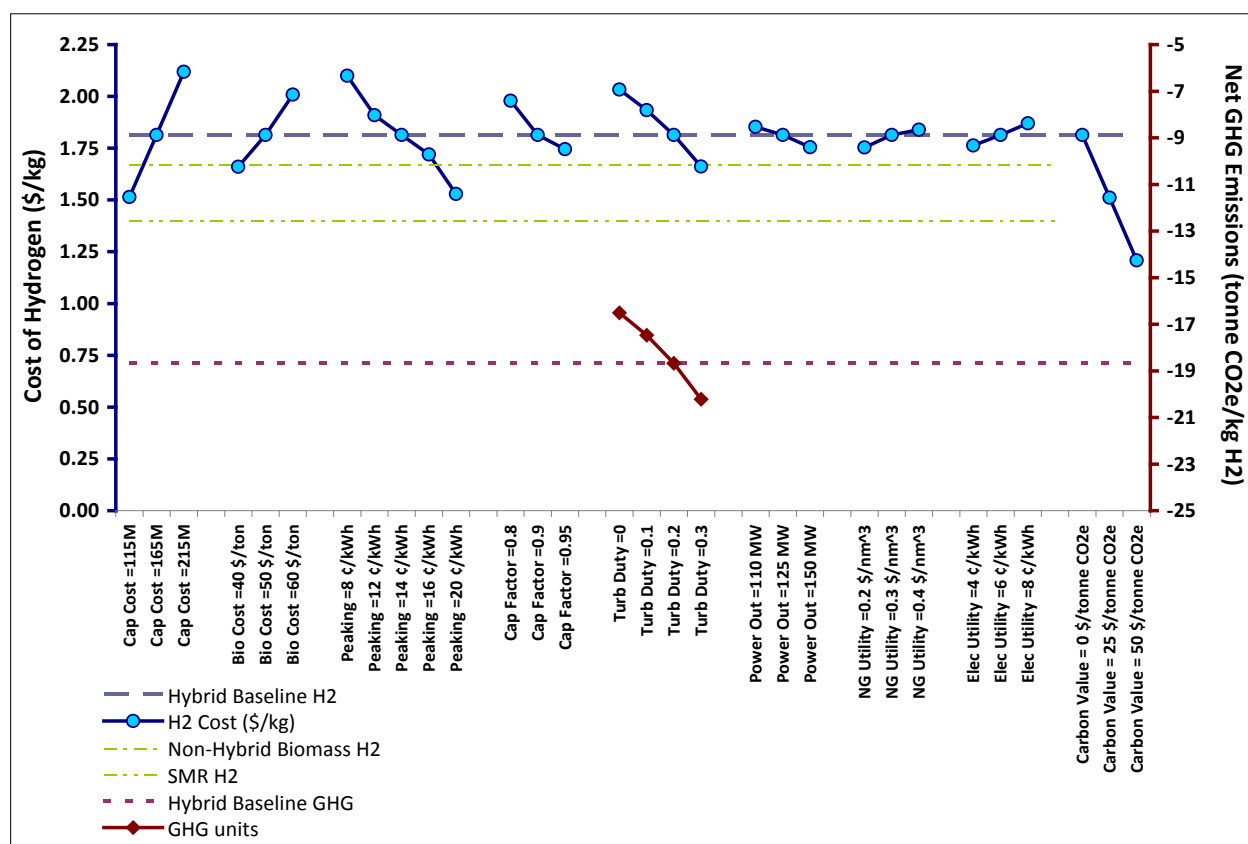


Figure 27. Integrated combined-cycle peaking sensitivity analysis, Northeast ISO

Indirect Hybrid System—Sinking

The proposed system switches between hydrogen production with syngas recycling for gasifier heat, and hydrogen production with electrical heating (sinking) to decrease or fully replace the syngas recycle. This sinking ability can best be described as a dispatchable load or demand from the viewpoint of the grid. Based on discussions with Xcel Energy, wind-generated electricity costs oftentimes can be negative, and having a dispatchable load would provide a valuable service to the utility.

No similar system was found to use for comparison, therefore a sinking duty of 20% was used as a starting point for the analysis. Table 19 summarizes the major economic model inputs by region. Unlike peaking electricity for which a contract rate was used for electricity produced by the plant, the sinking analysis simply used the average cost of electricity for the cheapest 20% of hours as the sinking electricity cost. The utility electricity cost was the average of the remaining 80% of the hours. These costs of electricity would be valid if a plant operator used accurate day-ahead energy market forecasts to schedule plant operation.

Table 19. Indirect Hybrid System—Sinking H2A Inputs

	NE ISO	MISO	NW Interface
Sinking duty	20%	20%	20%
Sinking electricity cost (¢/kWh)	4.35	2.20	2.18
Utility electricity cost (¢/kWh)	7.05	4.87	4.69
Utility natural-gas cost (\$/nm ³)	0.32	0.32	0.32
Cost of biomass (\$/ton)	48.83	48.83	48.83

Given the economic inputs above, and the plant inputs and outputs previously discussed, the cost of hydrogen production in each area was calculated. The results are shown in Table 20 along with the cost of hydrogen production for a baseline, non-hybridized biomass-to-hydrogen plant and a steam methane reforming plant.

Table 20. Indirect Hybrid System—Sinking Results

	\$/kg H₂ NE ISO	\$/kg H₂ MISO	\$/kg H₂ NW Interface
SMR	1.40	1.40	1.40
Biomass-to-hydrogen baseline	1.64	1.64	1.64
Electric air heater hybrid system	1.75	1.67	1.68
All electric heat hybrid system	1.77	1.68	1.67

There is a \$0.03 to \$0.13 premium on hydrogen produced by the proposed hybrid system compared with a non-hybrid biomass-to-hydrogen gasification plant. The additional capital costs of the sinking hybridization are not fully offset by additional revenue in any of the locations studied. The marginal costs found, however, are small enough that it is difficult to draw any definitive conclusion. Areas with lower-cost electricity move this hybridization closer to economic feasibility with the best results found in the Northwest.

Because the premium is small (about 5%) it might be acceptable in the long term. Recent studies have shown that there is inherent value added—or welfare effects—for electricity storage capacity (Sioshansi et al. 2009). Whether similar value is added by the proposed sinking hybrid is unknown. A more likely parallel would be the idea of “interruptible customers” which get discounted electricity rates in return for intermittent power supply. A similar contractual agreement could be envisioned for the proposed hybrid, in which discounted electricity rates would be provided in return for intermittent usage.

The proposed system is not a direct competitor with storage systems such as pumped hydro or compressed air energy storage (CAES). Energy storage systems attempt to profit by market arbitrage (selling electricity back to the grid at a price greater than that at which it was bought) whereas the proposed hybrid system sinks cheap electricity into transportation fuel. This could be considered cross-market arbitrage and the most similar system to this would be electrolysis. Compared to electrolysis, the proposed system is significantly less expensive and has the added benefit of running without electric heat when electricity costs are too high.¹¹

Assuming that all sinking electricity is renewable and taking the additional differences in carbon emissions into account, a value of \$26 to \$35 per metric ton of CO₂e makes the proposed hybrid cost competitive depending on the location and amount of dispatchable demand of the system. At \$27 per metric ton of CO₂e, the combustion air heater hybridization becomes cost competitive with a methane steam reforming plant in the Northwest. At \$26 per metric ton of CO₂e, the all-electric heat system becomes cost competitive with SMR in the Northwest.

¹¹ Based on \$3.24/kg hydrogen as reported in the “Future Central Hydrogen Production from Grid Electrolysis version 2.1.1” H2A analysis. Available at http://www.hydrogen.energy.gov/h2a_prod_studies.html. Accessed November 8, 2009.

Again, the baseline biomass-to-hydrogen plant only requires approximately \$23 to \$25 per metric ton of CO₂e value to be cost competitive with hydrogen produced by SMR. This means that peaking hybridization will be economically promising only when there is some value placed on the additional functionality of dual-mode operation.

To characterize the effect of various technical and economic assumptions, a sensitivity analysis was performed for both the sinking systems. Figure 28 and Figure 29 show the results for both the price of hydrogen and greenhouse gas emissions. The sensitivity analysis for both heating systems showed that capital costs, the cost of biomass, and the plant capacity factor are key inputs. Capital costs, when varied +/- 30%, cause the cost of hydrogen to vary \$0.18. The cost of biomass was varied from \$40 per ton to \$60 per ton based on projections in the MYPP (based on HyARC Energy Constants and Assumptions that are part of the H2A Analysis Tool) and causes the cost of hydrogen to vary by less than \$0.15. Changes in the sinking duty, the added hydrogen production due to sinking, the cost of natural gas, and the cost of electricity cause hydrogen costs to vary by less than 3%.

From the sinking duty sensitivity analysis, it appears that an increased sinking duty would be preferable to the 20% assumption made. Minimum hydrogen production prices occur when the electric heating systems are run approximately 40% of the time. This increases plant hydrogen production and decreases plant GHG emissions.

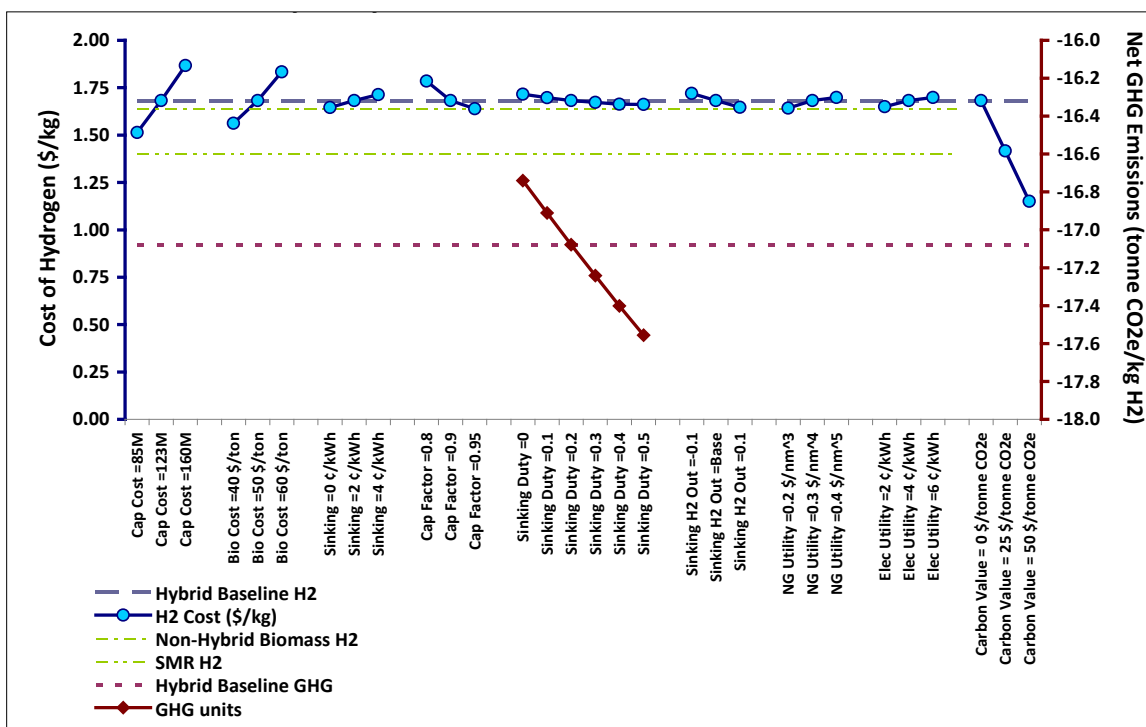


Figure 28. Air heater sinking sensitivity analysis, Northwest Interface

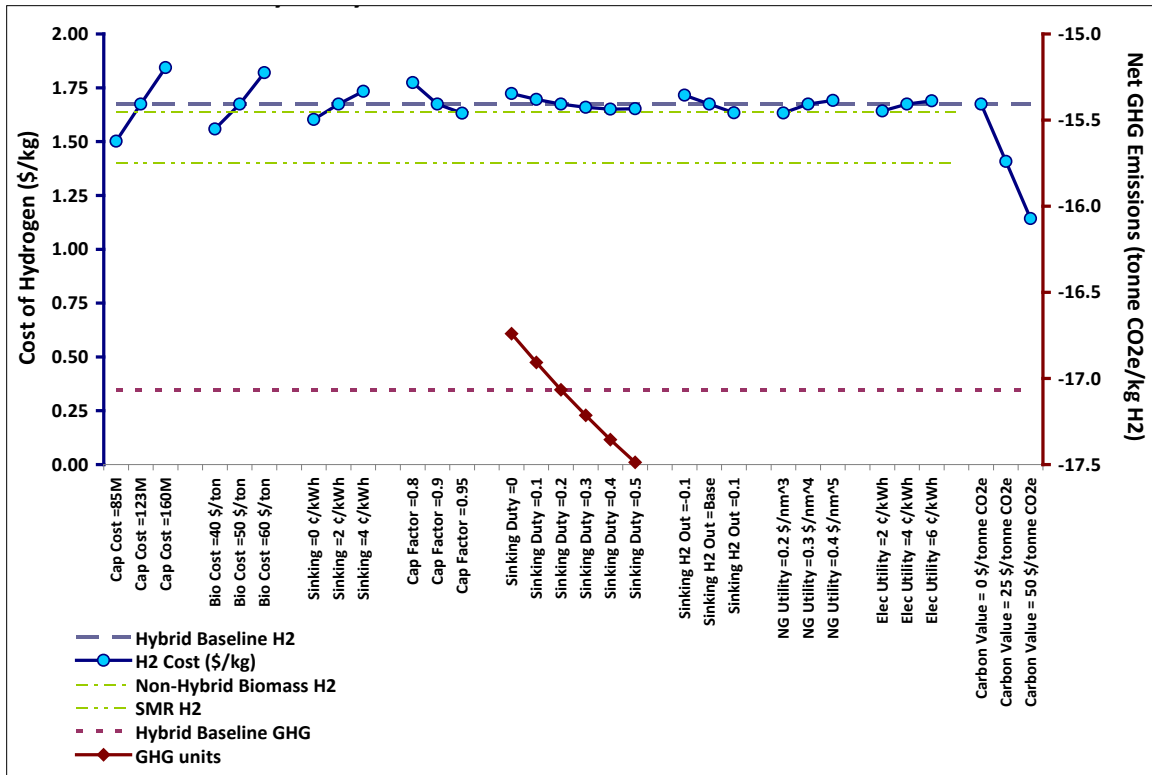


Figure 29. All-electric sinking sensitivity analysis, Northwest Interface

Direct Gasifier Hybrid System

Based on the previously described model inputs and outputs, an H2A analysis was completed to determine the baseline cost of hydrogen produced by a directly heated gasification plant that uses electrolysis regardless of electricity cost, and the cost of hydrogen for a directly heated gasification plant that uses a traditional ASU. Table 21 summarizes the major model. The plant was assumed to be located in the Midwest ISO region.

Table 21. Direct Hybrid System—H2A Inputs

	MISO
Electrolyzer duty (%)	100
Utility electricity cost (¢/kWh)	4.33
Cost of biomass (\$/ton)	48.83

The results of the analysis, along with the associated carbon emissions, are listed in Table 22. Costs for production of hydrogen via SMR and for an electrolysis plant only also are shown for reference purposes.

At \$2.32 per kilogram, hydrogen produced by a direct gasifier/electrolyzer hybrid plant is significantly more expensive than that produced by SMR. Electricity costs account for 36.9% of the overall cost of hydrogen produced (or \$0.86). Therefore, if electricity costs could be halved by intermittent operation, then the savings would bring the cost of hydrogen to \$1.89 per kilogram of hydrogen. At this price, the hybrid system could compete with hydrogen produced by a standard direct-gasification plant. It's important to note that these costs are approximately

15% greater than the \$1.64 per kilogram of hydrogen estimated for the indirectly heated baseline gasification plant used for comparison of the indirect hybrid analysis (*see* Table 20).

Table 22. Cost of Hydrogen and CO₂e Emissions from Direct Gasification

	\$/kg H₂	Net CO₂e/kg H₂
SMR	\$1.40	-6.8
Biomass gasification	\$2.13	-19.6
Electrolysis	\$2.59	14.5 (-17.8)
Electrolysis and gasification	\$2.32	0.24 (-18.2)

Unfortunately, to halve the cost of electricity used by the plant, the electrolyzer bank could be run only at the bottom of the area's price-duration curve. The previous sinking analysis assumed that extra electricity was used in only the cheapest 20% of hours in a year, which resulted in half-price sinking electricity. Intermittent operation significantly adds to plant capital costs. The electrolyzer bank must be increased in size drastically for intermittent operation so that it produces excess oxygen for storage. The additional electrolyzer costs, in addition to the added oxygen storage costs and operating expenses, make the potential electricity savings difficult to justify.

In addition to high production costs, carbon emissions actually increase relative to biomass gasification for this hybridization. If upstream emissions for electricity production are taken into account, electrolysis and this proposed hybrid are both net CO₂ emitters. This makes their justification by carbon value impossible unless only renewable electricity is used for operation (those values are shown in parentheses in Table 22). Since renewable wind energy is inherently intermittent, it is unreasonable to assume that renewable electricity is used without adding the necessary equipment for intermittent operation. One exception would be the use of hydroelectric power for the systems.

Similar to the other hybrid systems investigated, hybridization results in a price premium. There must be some additional justification for hybridization, such as welfare effects or utility demands. The baseline biomass-to-hydrogen case with an ASU requires only \$35 per metric ton of CO₂e to be cost competitive with SMR-produced hydrogen. Figure 30 shows the results of a sensitivity analysis performed on the baseline biomass to hydrogen via direct gasification model. Similar to the previous gasification systems, capital cost uncertainties have the greatest effect on the price of hydrogen produced. The sell price of electricity, electricity production, and water consumption have very little effect on overall economics of the plant.

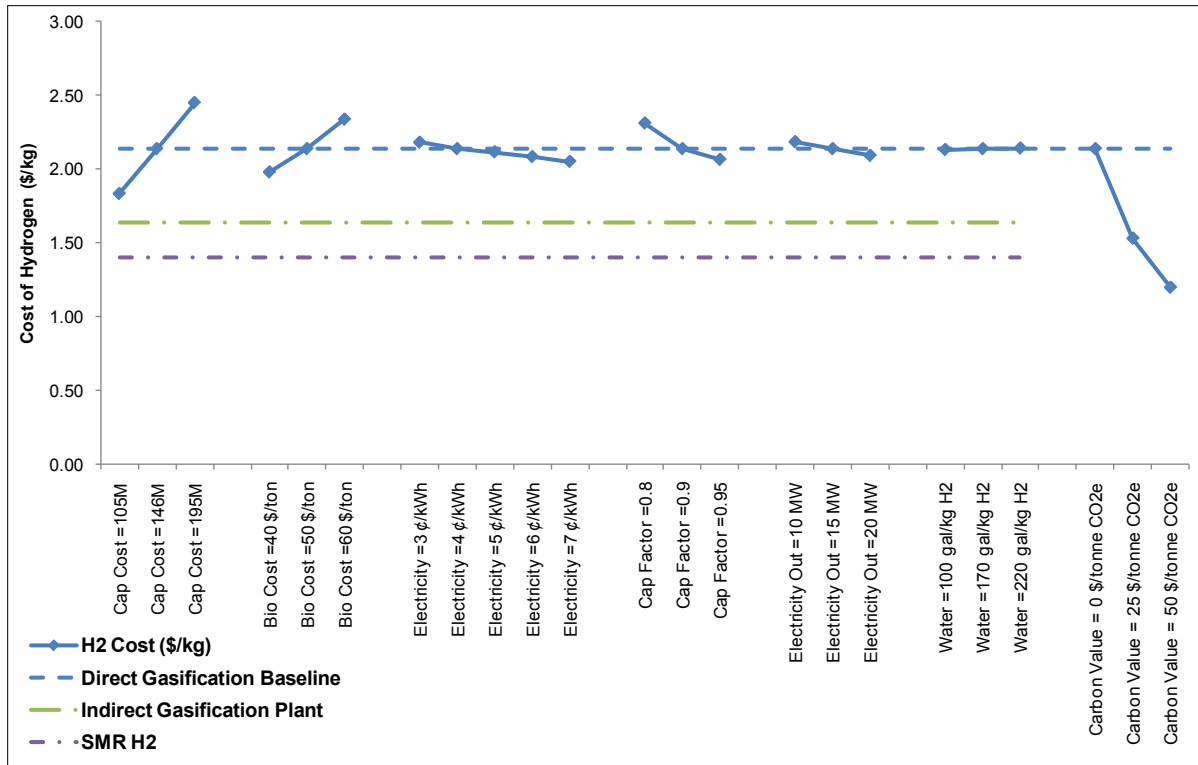


Figure 30. Directly heated gasifier baseline sensitivity analysis

In contrast to the baseline plant, electrolyzer-based hybrid system economics are extremely sensitive to the cost of electricity (*see* Figure 31). Fluctuations of plus or minus \$0.01 in the cost of electricity cause the price of hydrogen to fluctuate more than \$0.25 per kilogram. Lines are included for both electrolysis alone and the baseline biomass plant for comparison. If the price of electricity were to drop below \$0.04 per kilowatt-hour then the proposed hybrid system could economically compete with a non-hybrid biomass-to-hydrogen plant. This low price level is unlikely, however, and any energy-market fluctuations would have a dramatic effect on economic viability.

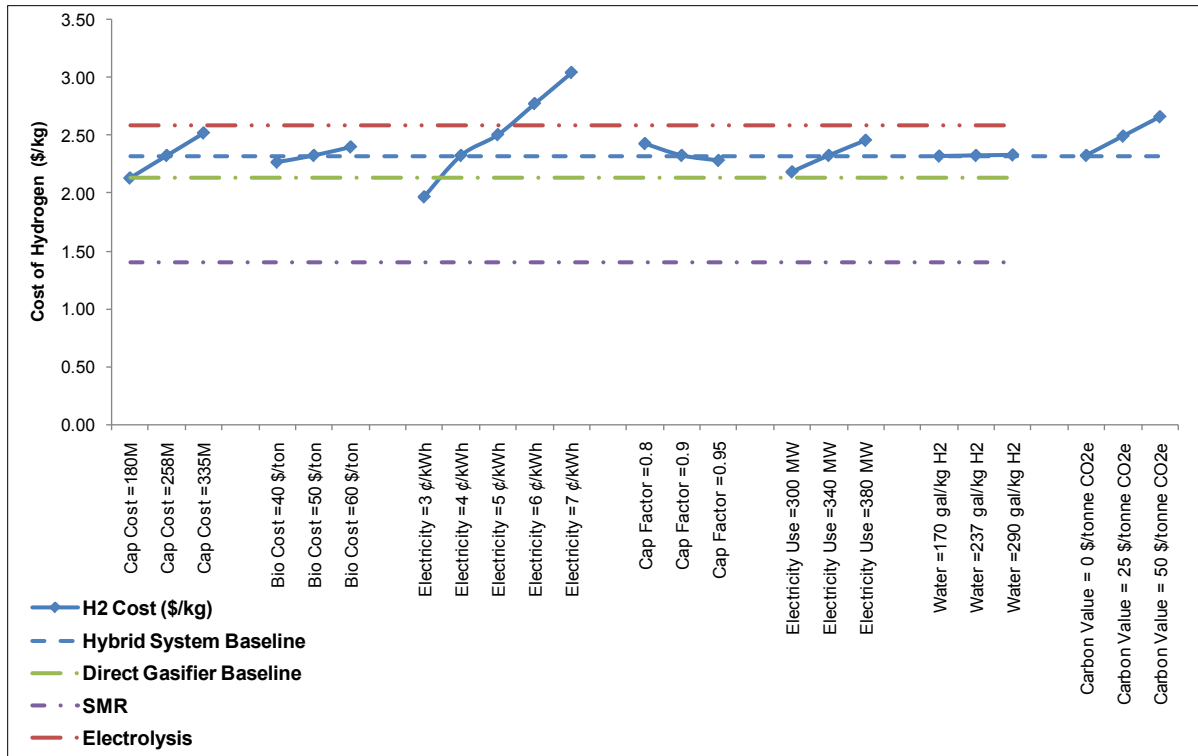


Figure 31. Directly heated gasifier hybrid system sensitivity analysis

Another biomass gasification, electrolysis hybridization for hydrogen production study was found after research began (Hutleberg and Karlsson 2009). The results of that study show a similar spread between gasification, electrolysis, and hybrid costs of hydrogen; however, the study was performed on a much smaller scale plant (approximately 37,000 kg/day H₂). Based on a conversion of the research results (in Swedish Krona) to dollars, the prices for hydrogen by gasification, electrolysis, and hybridization were \$5.55, \$6.15, and \$5.85 per kilogram of hydrogen produced, respectively.

Further intermittent simulation was not performed on the concept after these initial findings. For the use of electrolysis to make sense in this setting, there must be a significant benefit from the ability to operate intermittently. One envisioned possibility is to directly couple the plant with a stranded wind resource so that wind energy can be used to produce transportable fuel. This would be especially promising if the “stranded” wind resource was near a major trucking route (fuel demand). Fuel costs for such a system would be significantly greater than those of an ASU-based system, however, and only are justifiable if shipping of fuel to the site was difficult. Another option would be to use enriched air rather than pure oxygen for gasification. This would prevent the production of fuel but would be acceptable for power production and the electrolyzers could produce a slipstream of hydrogen fuel.

Conclusion

The indirect gasification concepts studied could be cost competitive in the near future as value is placed on controlling carbon emissions. Carbon values of slightly less than \$40 per metric ton of CO₂e make the systems studied cost competitive with steam methane reforming to produce

hydrogen. A non-hybrid biomass-to-hydrogen plant will be more cost competitive in general, however, so there must be some additional value placed on peaking or sinking for these plants to be economically attractive. This additional value is likely to become a reality as additional intermittent renewable energy sources—such as wind—are added to the national grid.

For the proposed peaking system, increased electricity production dramatically improves the economics so the additional complexity of a swinging integrated combined-cycle system could be justified. Further work is needed, however, to investigate the dynamic response and control of the proposed plants. Development of control strategies for these plants would be a significant undertaking and this topic was not addressed by the steady-state models used in this study.

The direct gasification concept studied is unlikely to be cost competitive in the near future. The additional electrolyzer costs make the possible electricity savings difficult to justify. Based on a direct replacement of the ASU with electrolyzers, hydrogen can be produced for \$2.32 per kilogram. Using grid electricity, however, the hybrid system is a net CO₂e emitter. For the use of electrolysis to make sense in this setting there must be a significant benefit to the ability to operate intermittently.

Future Work

As mentioned above, hybrid biomass systems encompass a wide range of possibilities. Significant opportunities exist for future work including the following.

- Determining whether there is inherent value added in electricity sinking/peaking that isn't directly considered by this analysis. Sioshansi et al. looked at the value of electricity storage in detail (Sioshansi et al. 2009) however; the market arbitrage they studied is significantly different from turning electricity into fuel. What is the value of “cross-market” arbitrage?

- Coal-biomass hybridization research received significant interest from multiple industry reviewers. Some research has been done on coal-biomass hybridization, as discussed in the literature review, but there are significant possibilities for future work including but not limited to the following.

 - A detailed techno-economic comparison of combined coal and biomass gasification systems versus coal alone or biomass alone.

 - Investigation of thermally integrated coal power plant and biomass/bio-oil gasification systems.

 - A comparison of the economic and welfare effects of the various methods for biomass and coal mixing including gasification, co-feed into steam boiler, bio-oil co-feed into steam boilers, and separate combustion.

- Industry reviewers expressed skepticism about the economic rationale for hydrogen production. Thus, an add-on study with economic comparison if liquid fuels such as Fischer-Tropsch fuels were produced instead of hydrogen would be informative.

The study focused specifically on near-term hybridization possibilities. A similar literature review and brainstorming process involving technologies on the horizon such as thermochemical hydrogen production cycles from water or oceanic biomass potential could yield interesting results.

The present study assumed biomass gasification plants that were optimized for hydrogen production and only producing electricity intermittently. Study of an inverse system in which a hydrogen slipstream is produced from a baseload power generation plant would yield significantly different results.

Price-duration curves were needed for each region studied. Predicting the effect that increased renewables on the grid would have on these curves could not be done with existing data. A study to create an “ideal” price-duration curve that could be used for similar studies would be extremely useful.

Further refinement is needed on the ICC electricity-production estimates and design because the ICC Aspen model was not a detailed plant model.

Determining how the system would behave dynamically when switched between modes of operation could be examined. Would the switch between hydrogen production and power production or electric heating and syngas recycling cause overall system instability? If the system is stable, how long would switching take and what are the thermal cycling effects?

Development of control strategies for each system could be studied. This can be done only after dynamic response is determined and is especially significant for the proposed integrated combined-cycle peaking concept.

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Appendix A. Numerical Idea Matrix

Analysis of Hybrid Hydrogen Production System Brainstorming									
Matrix Description	Initial decision matrix used to identify possible synergies between biomass processes and other existing and/or renewable processes. Zero represents no or limited correlations seen; 9 represents significant input/output correlation between the two technologies.								
Created	7/24/08				Researcher		Jered Dean		
Updated	11/22/08				Principle Investigator		Michael Penev		
INPUT VALUES FOR RANKING:	PAF	RPF	GHE	TRF	Biomass Gasification	Biomass Pyrolysis	Bio-oil Reforming	Biomass Combustion	Biomass Fermentation
PAF					10	10	10	10	2
PV Solar	7	9	9	7	3	3	1	0	0
Concentrated Solar	5	9	9	7	9	9	9	3	9
Thermal Photovoltaic	2	9	9	3	3	3	0	3	0
Photo-electrochemical Solar	2	9	9	3	1	1	0	0	0
Photo-bio Electrolysis	2	9	9	1	3	1	0	1	3
Wind Turbine (electrical)	9	9	9	9	3	3	1	0	0
Wind Turbine (mechanical)	7	9	9	9	1	1	0	1	3
Thermal Integration w/ Nuclear	4	5	9	7	9	1	9	0	3
Coal/Biomass Co-feed	9	5	7	7	9	3	3	9	1
Natural Gas Turbine	9	5	7	9	9	3	9	3	1
Steam Turbine	9	5	9	9	9	3	9	9	0
Thermal Integration w/ Coal	9	5	7	7	3	3	9	0	3
Alkaline Fuel Cell	7	5	9	3	0	0	0	0	0
PEM Fuel Cell	5	5	9	5	0	0	0	0	0
Phosphoric Acid Fuel Cell	2	5	9	7	0	0	0	0	0
SOFC/MCFC	7	5	7	7	9	3	9	0	3
Hydro-electric	3	9	9	9	0	0	0	0	0
PEM Electrolysis	2	5	9	5	9	0	9	0	0
Alkaline KOH Electrolysis	3	5	9	7	3	0	3	0	0
Hybrid SOFC/SOFC Electrolysis	2	5	7	3	9	3	9	1	0
Solid Oxide Electrolysis	3	5	7	3	9	0	9	0	0
Thermo-chemical cycles	3	5	9	3	3	1	1	3	0
Wastewater Treatment	9	5	9	9	9	9	3	3	9
Notes: Weighting Factors are Potential Availability Factor (PAF) and Renewable Potential Factor (RPF) and Greenhouse Gas Emissions (GHE). The higher the PAF the more readily available plant and fuel are for the process. The RPF was divided into non-renewable (0), partially renewable or dependant on operation parameters (5) and 100% renewable (9). GHE was scored as no direct emissions (9), emissions that could be sequestered or reprocessed (7), and low emissions (5). Technology Readiness Factor (TRF)									

Appendix B. Indirect Gasifier Aspen Plus Simulation Details

All Aspen Plus simulations were modifications to the previous Wood to Hydrogen Using Indirectly Heated Gasifier model available at http://devafdc.nrel.gov/biogeneral/Aspen_Models/. Specifically, the design for “goal” tar reforming technology was used. Every attempt was made to maintain the base model thermal integration and minimize changes to the model. Details of the modifications made are given below.

GE 6FA Simple Cycle Gas Turbine

The following modifications were made to the existing Wood to Hydrogen Aspen model.

The natural gas trim to the tar reformer catalyst regenerator (stream 427) flow was set to zero and the NGTRIM design spec was deactivated.

The PSA system and all downstream flows for hydrogen compression were set to zero and deactivated.

Stream 420 that normally would flow through the PSA system instead was routed to the tar reformer catalyst regenerator R-204.

The gas-turbine model developed separately was inserted into the model.

Syngas stream 326 (directly after the LO-CAT reactor) was split to the water-gas shift reactors and the turbine respectively. Any remaining syngas was considered waste. The following constraints were met with design specifications.

TARFEED—Adjusted the split fraction until the difference between the catalyst regenerator and the tar reformer was 200 °F.

TBFEED—Adjusted the split fraction until 6FA combustor temperature reached the specified 1,288 °C.

A heat exchanger was added to the turbine exhaust so that lost heat to the plant steam turbine from H-405 and H-407 heat exchangers due to less flow through the water-gas shift is exactly replaced by the turbine exhaust. This was controlled with calculator block “THEAT.”

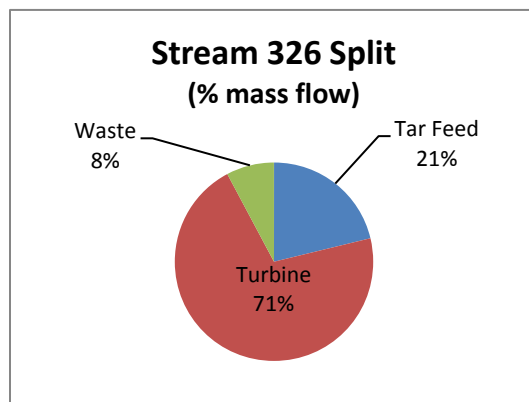


Figure B.1. Syngas stream 326 split

A screen capture of the gas turbine inputs and outputs is provided as Figure B.2, below.

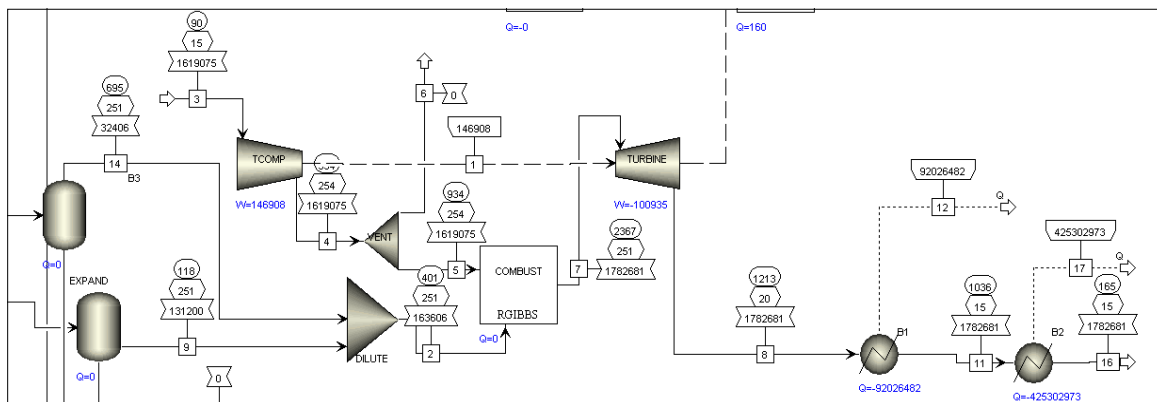


Figure B.2. Gas turbine inputs and outputs

The final plant power specifications for this scenario are summarized in Table B.1.

Table B.1. Plant Power Generation

	HP	MW
WPLANT	47,442	35.4
WGEN	-43,154	-32.2
WTURB	-100,935	-75.3
WNET	-96,647	-72.1

GE F-Class Simple-Cycle Gas Turbine

The following modifications were made to the case v024 Aspen model.

Syngas stream 326 (directly after the LO-CAT reactor) was split to the water-gas shift reactors and the turbine respectively. The following constraints were met with design specifications.

TARFEED—Adjusted the split fraction until the difference between the catalyst regenerator and the tar reformer was 200°F. Stated another way, stream 107 was maintained at approximately 1,791°F.

TBFEED—All but available syngas remaining after the TARFEED design specification was met was sent to the gas turbine. Additional calculators were added so that the air and steam to fuel ratios remained constant with the excess fuel flow.

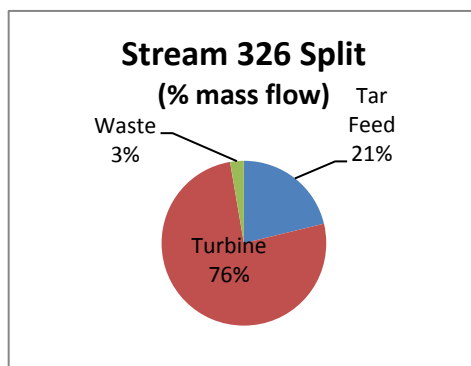


Figure B.3. Syngas stream 326 split

Please note that 5,000 lb/hr of the syngas stream still was sent to waste in this case. The Biomass to Hydrogen model recently was updated to include new yield correlations for the gasifier, based on data collected in the Thermochemical Process Development Unit (Kinchin and Bain 2009). The updated model requires raw syngas to be diverted and combusted to supplement the heat delivered to the gasifier by the char combustor. The 5,000 lb/hr waste stream effectively adjusts the energy flow in the current model to align with (Kinchin and Bain 2009) findings. A screen capture of the gas turbine inputs and outputs is supplied as Figure B.4, below.

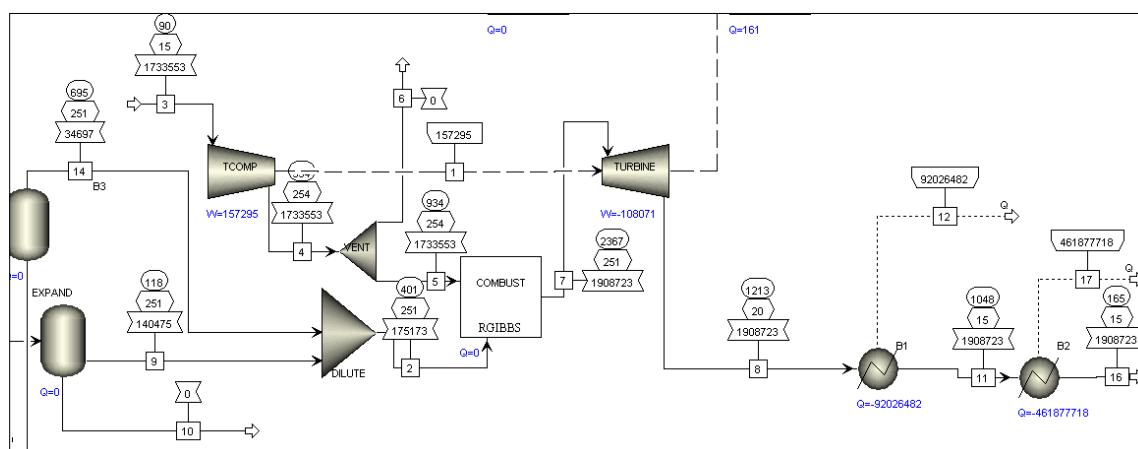


Figure B.4. Gas turbine inputs and outputs

The final plant power specifications for this scenario are summarized in Table B.2.

Table B.2. Plant Power Generation

	HP	MW
WPLANT	47,435	35.4
WGEN	-43,163	-32.2
WTURB	-108,071	-80.6
WNET	-103,799	-77.4

Water-gas Shift Shutdown

The following modifications were made to the case v025 Aspen model.

Syngas stream 326 (directly after the LO-CAT reactor) was split to the tar reformer catalyst regenerator and the turbine respectively. The following constraints were met with design specifications.

TARFEED—Adjusted the split fraction until the difference between the catalyst regenerator and the tar reformer was 200°F.

TBFEED—All but available syngas remaining after the TARFEED design specification was met was sent to the gas turbine. Additional calculators were added so that the air and steam to fuel ratios remained constant with the excess fuel flow.

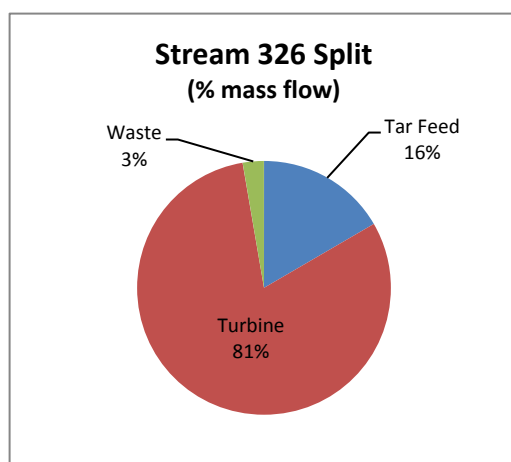


Figure B.5. Syngas stream 326 split

Please note that 2,268 kg/hr (5,000 lb/hr) of the syngas stream still was sent to waste in this case. The Biomass to Hydrogen model recently was updated to include new yield correlations for the gasifier, based on data collected in the Thermochemical Process Development Unit (Kinchin and Bain 2009). The updated model requires raw syngas to be diverted and combusted to supplement the heat delivered to the gasifier by the char combustor. The 2,268 kg/hr waste stream effectively adjusts the energy flow in the current model to align with (Kinchin and Bain 2009). The final plant power specifications for this scenario are summarized in Table B.3.

Table B.3. Plant Power Generation

	HP	MW
WPLANT	47,435	35.4
WGEN	-43,163	-32.2
WTURB	-114,652	-85.5
WNET	-110,380	-82.3

Appendix C. Aspen Plus Gas Turbine Model Details

Table C.1. GE Gas Turbine Specifications

	MS6001FA
Output (MW)	75.9 (90)
Heat rate (kJ/kWh)	10,332
Pressure ratio	15.7:1
Mass flow (kg/sec)	204
Turbine speed (rpm)	5,254
Exhaust temp (°C)	603
Turbine inlet temp (°C)	1,288

Based on specifications from GE (summarized in Table C.1), a simple Aspen Plus model was constructed of the 6FA simple-cycle gas turbine. The flow sheet is shown in Figure C.1 below.

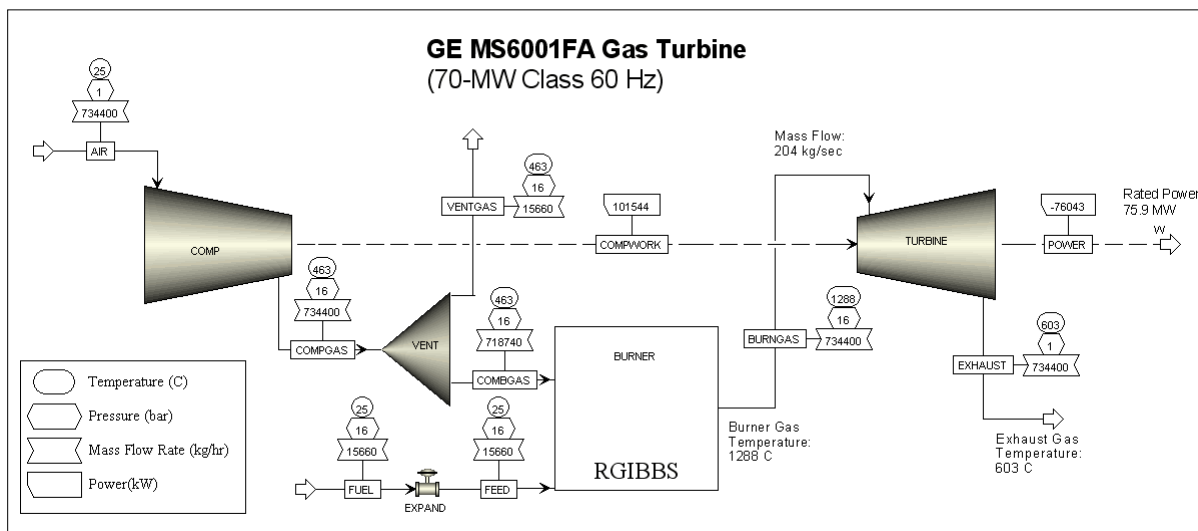


Figure C.1. Simple gas turbine flow sheet

The model was calibrated in the following manner.

The heat rate was set so that 10,332 kJ/kWh of methane (lower heating value) to the RGIBBS burner at ambient temperature and 15.7 bar pressure.

The mass flow through the compressor was set to 204 kg/sec of air with a pressure ratio of 15.7.

A calculator block was specified that sets the total “BURNGAS” flow to 204 kg/sec (vents compressor gas as needed).

The compressor isentropic efficiency was adjusted until burner temperature reached 1,288°C.

The turbine isentropic efficiency was adjusted until the outlet temperature was 603°C.

The compressor mechanical efficiency was adjusted until the power output was approximately 75.9 MW.

Table C.2. Aspen Plus Natural Gas Model Specifications

	Compressor	Turbine
Mech efficiency (%)	93	100
Isentropic process efficiency (%)	77.8	92.5
Compression ratio	15.7	–
Outlet pressure (psia)	–	14.7

Table C.2 summarizes the natural gas compressor and turbine settings used for all simulations. To run the turbine on hydrated syngas, the following additional steps were taken.

The feed was adjusted until the burner temperature reached the target temperature of 1,288°C.

The calculator block was modified so that up to 14% extra flow (more than the 204 kg/sec limit) could be passed through the turbine.

The compressor pressure ratio was adjusted so that the volumetric flow through the turbine remained the same as the natural gas basecase.

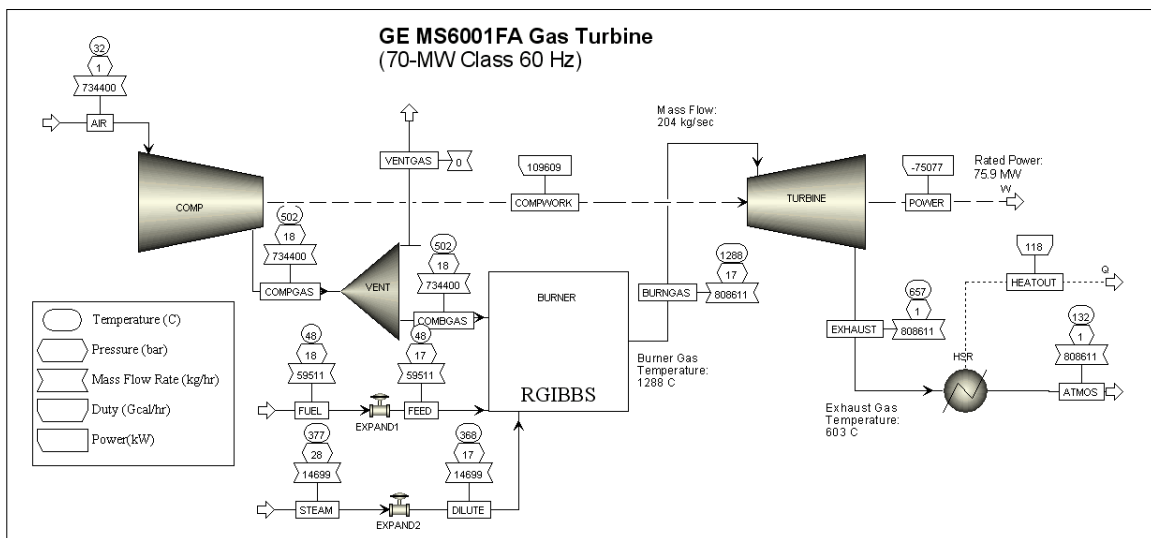


Figure C.2. Natural gas turbine flow sheet

To estimate the combined-cycle outputs the following additional steps were taken.

An additional heater block was inserted in the turbine exhaust stream that took the exhaust temperature to 132°C.

The heat duty of this heater block was taken times the efficiency of 0.40 to estimate the ideal steam cycle power available.

Appendix D. Electric Heater Capital Costs

Table D.1 presents the cost estimate for adding only electric combustion air preheating to the gasifier plant.

Table D.1. Cost Estimate

Cost of 2.2 MW unit today	\$	250,000.00
Size priced	kW	2,200.00
Electricity to heat efficiently	—	0.90
Size needed	kW	96,191.97
Scaling factor	—	0.9
Cost of 86.5-MW unit today	\$	7,491,771.63
Cost of 86.5-MW unit 2005	\$	6,396,147.84
Installation factor	—	2.47
Total 2005 installed cost	\$	15,798,485.16

Cost of base air blower	\$	34,860.00
Size priced	kW	1,857.54
Size needed	kW	3,885.10
Scaling factor	—	0.6
Cost of needed unit today	\$	54,275.74
Cost of needed unit 2005	\$	46,338.26
Installation factor	—	2.47
Total 2005 installed cost	\$	114,455.51
Basecase cost	\$	91,105.77
Incremental increase	\$	23,349.74

Total capital		\$15,821,834.90
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Table D.2 provides the capital cost estimate for completely replacing syngas recycle stream with electric heat.

Table D.2. Cost Estimate

Cost of 2.2 MW unit today	\$	250,000.00
Size priced	kW	2,200.00
Size needed	kW	86,572.77
Scaling factor	—	0.9
Cost of 86.5 MW unit 2009	\$	6,814,010.35
Cost of 86.5 MW unit 2005	\$	5,817,504.82
Installation factor	—	2.47
Total 2005 installed cost of air heater	\$	14,369,236.91
Cost of base air blower	\$	34,860.00
Size priced	kW	1,857.54
Size needed	kW	5,827.64
Scaling factor	—	0.6
Cost of needed unit today	\$	69,224.61
Cost of needed unit 2005	\$	59,100.96
Installation factor	—	2.47
Total 2005 installed cost	\$	145,979.36
Basecase cost	\$	91,105.98
Incremental increase	\$	54,873.38
Inconel heat flux capacity	W/cm ²	13.00
Size needed	kW	23,632.72
Area needed	cm ²	1,817,901.56
Diameter of rod	cm	1.59
Length of rod needed	m	3,639.45
Cost of rod (2008)	\$/m	175.43
Total material cost today	\$	638,476.92
Cost of heater factor	—	2.50
Total cost of heater today	\$	1,596,192.31
Cost of heater 2005	\$	1,362,759.37
Installation factor	—	2.47
Total 2005 installed cost	\$	3,366,015.65
Total capital range for modification		\$17,790,125.93

Appendix E. Direct Gasifier Baseline Capital Costs

	Number Required	Number Spares	Equipment Name	Units	Scaling Stream Flow	Original Stream Flow	Size Ratio	Original Equipment Cost (per Unit)	Base Year	Total Original Equip Cost (Req'd & Spare)	Scaling Exponent	Scaled Cost in Base Year	Installation Factor	Installed Cost in Base Year	Installed Cost in 2005\$	Subtotals	Source
Feed Prep	2		Feed Prep	wet tonne/hr	83	65	1.29	\$ 9,840,000	2003	\$ 19,680,000	0.77	\$ 23,935,511	1	\$ 23,935,511	\$ 27,897,950	\$ 27,897,950	b
	2		Gasifier	dry tonne/hr	73	42	1.76	\$ 6,410,000	2003	\$ 12,820,000	0.7	\$ 19,026,792	1	\$ 19,026,792	\$ 22,176,609	\$ 22,176,609	b
	2		Primary Cyclone	actual m3/s gas	10	69	0.14	\$ 910,000	2003	\$ 1,820,000	0.7	\$ 469,033	1	\$ 469,033	\$ 546,680	\$ 22,723,289	b
ASU	1		ASU	tonne/hr pure O2	28	77	0.36	\$ 22,700,000	2003	\$ 22,700,000	0.5	\$ 13,680,124	1	\$ 13,680,124	\$ 15,944,820	\$ 15,944,820	b
	1		N2 Compressor	MWe consumed	2	10	0.20	\$ 4,140,000	2003	\$ 4,140,000	0.67	\$ 1,402,618	1	\$ 1,402,618	\$ 1,634,816	\$ 1,634,816	b
	1		O2 Compressor	MWe consumed	4	10	0.45	\$ 5,540,000	2003	\$ 5,540,000	0.67	\$ 3,225,740	1	\$ 3,225,740	\$ 3,759,750	\$ 21,339,385	b
Syngas Cleaning	1		Syngas Cooler	kW	60,000	77,000	0.78	\$ 25,400,000	2003	\$ 25,400,000	0.6	\$ 21,869,056	1	\$ 21,869,056	\$ 25,489,400	\$ 25,489,400	b
	1	1	Sludge Pump	kg/hr	500	452	1.11	\$ 3,911	2002	\$ 7,822	0.33	\$ 8,084	2.47	\$ 19,968	\$ 23,633	\$ 23,633	a
	1		Sludge Settling Tank	kg/hr	3,706	3,706	1.00	\$ 11,677	2002	\$ 11,677	0.6	\$ 11,677	2.47	\$ 28,842	\$ 34,135	\$ 34,135	a
Fuel Synthesis	1		LO-CAT Oxidizer	kg/hr	360	234	1.54	\$ 1,000,000	2002	\$ 1,000,000	0.65	\$ 1,324,631	2.47	\$ 3,271,837	\$ 3,872,281	\$ 3,872,281	a
	1		Vessel	kW	18,189	15,121	1.20	\$ 71,389	2002	\$ 71,389	0.44	\$ 77,435	2.47	\$ 191,263	\$ 226,364	\$ 226,364	a
	1		ZnO Bed Preheater	kg/hr	117,088	83,771	1.40	\$ 37,003	2002	\$ 74,006	0.56	\$ 89,269	2.47	\$ 220,494	\$ 260,959	\$ 29,906,771	a
	2		ZnO Sulfur Removal Beds	kg/hr													
Fuel Synthesis	1		High Temperature Shift Reactor	kg/hr	161,102	146,450	1.10	\$ 465,907	2002	\$ 465,907	0.56	\$ 491,461	2.47	\$ 1,213,908	\$ 1,436,683	\$ 1,436,683	a
	1		LT Shift														
	1		Precooler/BFW Preheater #1	kW	17,523	25,035	0.70	\$ 56,089	2002	\$ 56,089	0.6	\$ 45,281	2.47	\$ 111,843	\$ 132,369	\$ 132,369	a
Fuel Synthesis	1		Low Temperature Shift Reactor	kg/hr	161,102	146,451	1.10	\$ 323,464	2002	\$ 323,464	0.56	\$ 341,204	2.47	\$ 842,774	\$ 997,439	\$ 997,439	a
	1		PSA														
	1		Precooler/Deaerator Water Preheater #2	kW	9,793	9,207	1.06	\$ 21,089	2002	\$ 21,089	0.6	\$ 21,885	2.47	\$ 54,055	\$ 63,975	\$ 63,975	a
Fuel Synthesis	1		PSA Air-cooled Precooler	kW	25,785	31,283	0.82	\$ 388,064	2002	\$ 388,064	0.6	\$ 345,572	2.47	\$ 853,563	\$ 1,010,207	\$ 1,010,207	a
	1		Pre-PSA Knock-out #1	kg/hr	161,102	146,451	1.10	\$ 129,979	2002	\$ 129,979	0.6	\$ 137,632	2.47	\$ 339,950	\$ 402,337	\$ 402,337	a
	1		Pre-PSA Knock-out #2	kg/hr	135,940	111,591	1.22	\$ 55,291	2002	\$ 55,291	0.6	\$ 62,242	2.47	\$ 153,738	\$ 181,952	\$ 181,952	a
	1		Pressure Swing Adsorption Unit	kg/hr	7,120	6,950	1.02	\$ 4,855,471	2002	\$ 4,855,471	0.6	\$ 4,926,334	2.47	\$ 12,168,045	\$ 14,401,109	\$ 18,626,072	a
Steam Cycle	1		Based on Installed Cost of 474.34 \$/kWh	kWh													
	1				36,380	29,973	1.21	\$ 14,217,400	2002	\$ 14,217,400	1	\$ 17,256,489	1	\$ 17,256,489	\$ 20,423,378	\$ 20,423,378	a
Subtotal																	\$ 140,916,844
Contingency (1.5%)																	\$ 2,113,753
Structures																	\$ 6,368,900
Total																	\$ 149,399,497

Spath et al. "Biomass to Hydrogen Production Detailed Design and Economics Utilizing the Battelle Columbus Laboratory Indirectly-Heated Gasifier". NREL/TP-310-37408. May 2005.

Sources: a Spath et. al. "Biomass to Hydrogen Production Detailed Design and Economics Utilizing the Battelle Columbus Laboratory Indirectly-Heated Gasifier", NREL/TP-510-37408, May 2005.
b Jin, H. et al. "Performance and cost analysis of future, commercially mature gasification-based electric power generation from switchgrass", BioPr Vol 3 pp 142-173, 2009.

Appendix F. Direct Gasifier Hybrid System Capital Costs

		Number Required	Number Spares	Equipment Name	Scaling Stream Units	Scaling Stream Flow	Original Stream Flow	Size	Original Equipment Base Cost (\$)	Total Original Equip Cost (Feed & Spare) in Base Cost (\$)	Scaling Extrapolation Factor	Scaled Cost in Base Cost (\$)	Installation Factor	Installed Cost in Base Cost (\$)	Installed Cost in Source (\$)	
Feed Prep	Feed Prep	2	2	Feed Prep	wet tonne/hr	83	65	1.29	\$ 9,840,000	\$ 19,680,000	0.77	\$ 23,935,511	1	\$ 23,935,511	\$ 27,897,950	Source b
		2	2	Feed Prep	wet tonne/hr	83	65	1.29	\$ 9,840,000	\$ 19,680,000	0.77	\$ 23,935,511	1	\$ 23,935,511	\$ 27,897,950	Source b
Gasifier	Gasifier	2	2	Gasifier	dry tonne/hr	73	42	1.76	\$ 6,410,000	\$ 12,820,000	0.7	\$ 19,026,792	1	\$ 19,026,792	\$ 22,176,609	b
		2	2	Primary Cyclone	actual m3/s gas	10	69	0.14	\$ 910,000	\$ 1,820,000	0.7	\$ 469,033	1	\$ 469,033	\$ 546,680	\$ 22,723,289
Electrolyzer	Electrolyzer	1	1	Electrolyzer Plant	Future H2A	4	10	0.45	\$ 5,540,000	\$ 5,540,000	0.67	\$ 3,225,740	1	\$ 3,225,740	\$ 95,402,426	c
		1	1	O2 Compressor	Mwe consumed	4	10	0.45	\$ 5,540,000	\$ 5,540,000	0.67	\$ 3,225,740	1	\$ 3,225,740	\$ 3,759,750	\$ 99,162,176
Syngas Cleaning	Syngas Cleaning	1	1	Syngas Cooler	kW	60,000	77,000	0.78	\$ 25,400,000	\$ 25,400,000	0.6	\$ 21,869,056	1	\$ 21,869,056	\$ 25,489,400	b
		1	1	Sludge Pump	kg/hr	300	452	1.11	\$ 3,911	\$ 7,822	0.33	\$ 8,084	2.47	\$ 19,988	\$ 23,633	a
		1	1	Sludge Settling Tank	kg/hr	3,708	3,708	0.00	\$ 11,677	\$ 11,677	0.6	\$ 11,677	2.47	\$ 28,842	\$ 34,135	a
		1	1	2-Stage Sulfuric Acid	Mwe consumed	2	10	0.20	\$ 4,140,000	\$ 4,140,000	0.67	\$ 1,402,618	1	\$ 1,402,618	\$ 1,659,916	b
		1	1	Gas Removal	kg/hr	117,280	310,258	0.38	\$ 59,505,000	\$ 59,505,000	0.6	\$ 33,193,515	1	\$ 33,193,515	\$ 31,107,293	d
		1	1	Elemental Sulfur Recovery	kg/hr	23	5,676	0.00	\$ 11,997,000	\$ 11,997,000	0.6	\$ 440,234	1	\$ 440,234	\$ 412,565	\$ 58,701,843
Fuel Synthesis	Fuel Synthesis	1	1	High Temperature Shift Reactor	kg/hr	161,102	146,450	1.10	\$ 465,907	\$ 465,907	0.56	\$ 491,461	2.47	\$ 1,213,908	\$ 1,436,683	a
		1	1	LT Shift Precooler/BFW Preheater #1	kW	17,523	25,035	0.70	\$ 56,089	\$ 56,089	0.6	\$ 45,281	2.47	\$ 111,843	\$ 132,369	a
PSA	PSA	1	1	Low Temperature Shift Precooler / PSA Precooler #1	kg/hr	161,102	146,451	1.10	\$ 323,464	\$ 323,464	0.56	\$ 341,204	2.47	\$ 842,774	\$ 997,439	a
		1	1	Deaerator Water Preheater #2	kW	9,793	9,207	1.06	\$ 21,089	\$ 21,089	0.6	\$ 21,885	2.47	\$ 54,055	\$ 63,975	a
		1	1	PSA Air-cooled Precooler	kW	25,785	31,283	0.82	\$ 388,064	\$ 388,064	0.6	\$ 345,572	2.47	\$ 853,563	\$ 1,010,207	a
		1	1	Pre-PSA knock-out #1	kg/hr	161,102	146,451	1.10	\$ 129,979	\$ 129,979	0.6	\$ 137,632	2.47	\$ 339,950	\$ 402,337	a
		1	1	Pre-PSA knock-out #2	kg/hr	135,940	111,591	1.22	\$ 55,291	\$ 55,291	0.6	\$ 62,242	2.47	\$ 153,738	\$ 181,952	a
		1	1	Pressure Swing Adsorption Unit	kg/hr	7,120	6,950	1.02	\$ 4,855,471	\$ 4,855,471	0.6	\$ 4,926,334	2.47	\$ 12,168,040	\$ 14,401,109	\$ 16,626,072
Steam Cycle	Steam Cycle	1	1	Based on installed Cost of \$74.34 /kWh	kWh	36,380	29,973	1.21	\$ 14,217,400	\$ 14,217,400	1	\$ 17,256,489	1	\$ 17,256,489	\$ 20,423,378	a
		1	1	Subtotal											\$ 247,534,706	
Total	Total	1	1	Contingency (1.5%)											\$ 3,713,021	a
		1	1	Structures											\$ 6,368,900	a
		1	1	Total											\$ 257,616,627	
Source:	a	Spath et al., "Biomass to Hydrogen Production Detailed Design and Economics Utilizing the Battelle Columbus Laboratory Indirectly-Heated Gasifier", NREL/TP-510-37408, May 2005.														
	b	Jin, H. et al., "Performance and cost analysis of future, commercially mature gasification-based electric power generation from switchgrass", Biofuels Vol 3 pp 142-173, 2009.														
	c	Future Central Hydrolysis Production from Grid Electrollysis version 2.1.1 Available at: http://www.hydrogen.energy.gov/ira/prc_studies.html														

	Total
Späth et al. "Biomass to Hydrogen Production Detailed Design and Economics Utilizing the Battelle Columbus Laboratory Indirectly-Heated Gasifier". NREL/TP-510-37408. May 2005.	
Li, H. et al. "Performance and cost analysis of power generation based gasification-based electricity power generation from switchgrass". <i>Bldg Res J</i> 3 pp 142-173. 2009.	
"Cost and Performance Baseline for Fossil Energy Plants, Vol. 1: DOE/NETL-2007/281". http://www.hydrogen.energy.gov/H2a_top_studies.html	
"Cost and Performance Baseline for Fossil Energy Plants, Vol. 1: DOE/NETL-2007/281".	

Subtotal	\$ 247,534,706
Contingency (1.5%)	\$ 3,713,021
Structures	\$ 6,368,900
Total	\$ 257,616,627

Appendix G. Electricity Profiles

Northeast ISO

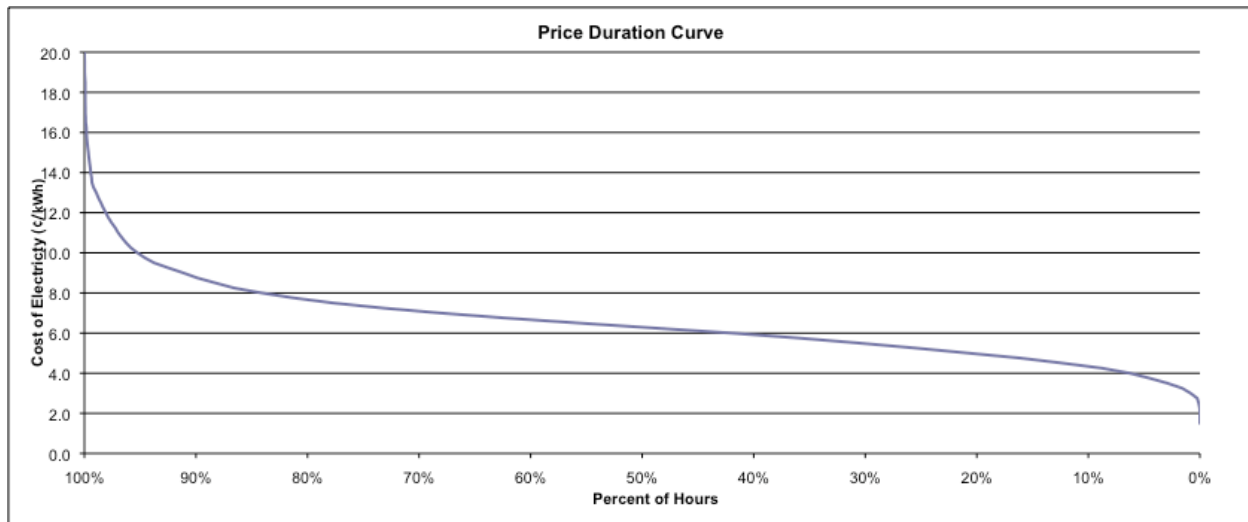


Figure G.1. Northeast ISO price duration curve

High average cost because of reliance on natural-gas turbines

Hourly day-ahead market data received directly from Northeast ISO Website

Average price of electricity is \$0.0655/kWh

Midwest ISO

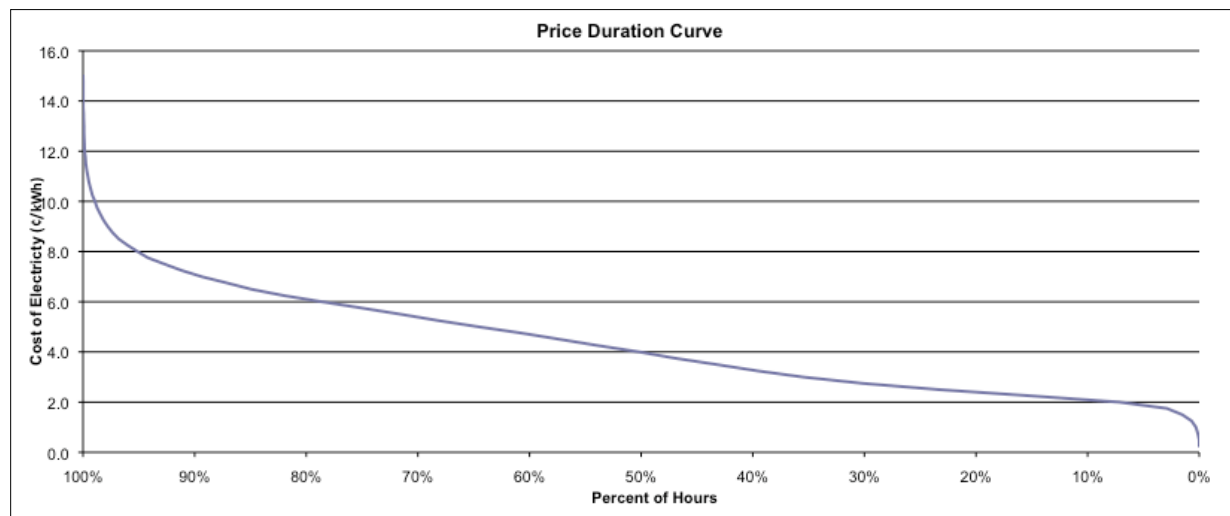


Figure G.2. Midwest ISO price duration curve

Low cost of electricity because of large amount of coal on the grid

Hourly day ahead market data received from MISO directly

Average price of electricity of \$0.0432/kWh

Northwest Interface

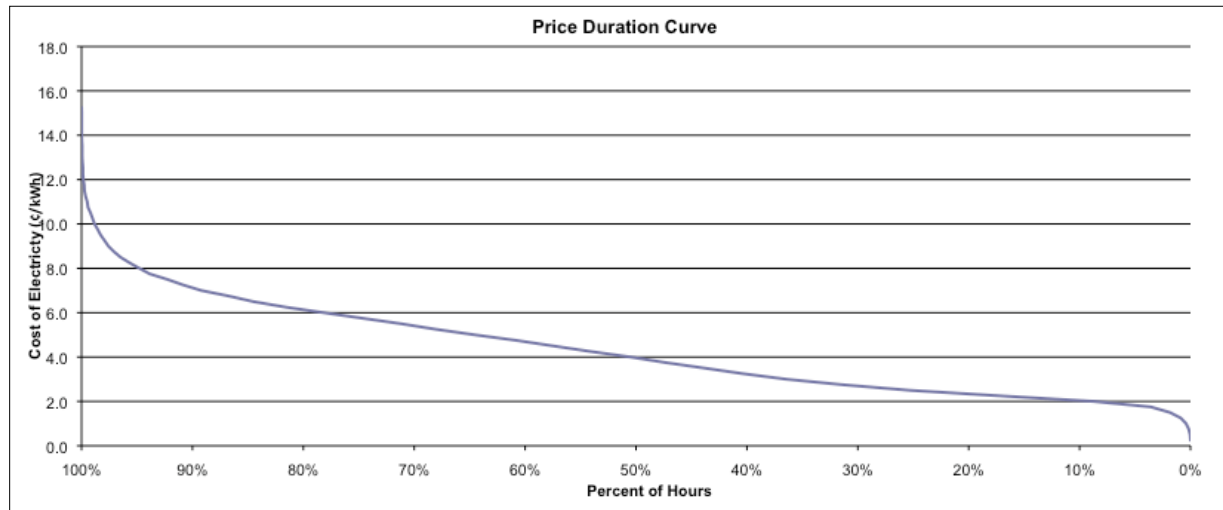
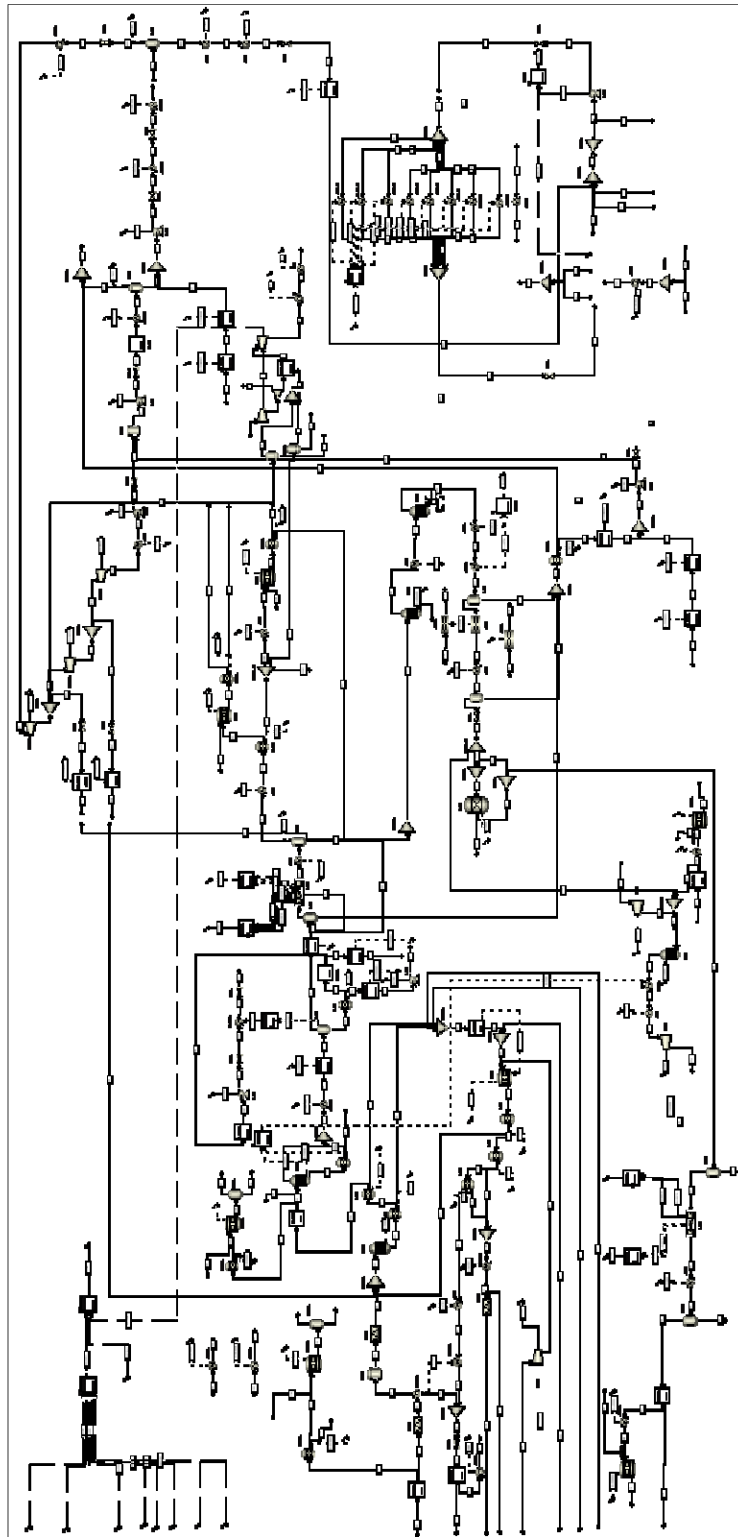


Figure G.3. Northwest Interface price duration curve

Low cost of electricity because of large amount of hydro on the grid
No RTO/ISO in the area, so the data from the northwest interface is used instead
Hourly load lambda data received from DOE records
Average cost of electricity of \$0.431/kWh

Appendix H. Aspen Flow Sheet



REPORT DOCUMENTATION PAGE

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				5b. GRANT NUMBER		
				5c. PROGRAM ELEMENT NUMBER		
6. AUTHOR(S) J. Dean, R. Braun, and D. Munoz: Colorado School of Mines M. Penev and C. Kinchin: NREL				5d. PROJECT NUMBER NREL/TP-560-46934		
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14. ABSTRACT (Maximum 200 Words) Report on biomass pathways for hydrogen production and how they can be hybridized to support renewable electricity generation. Two hybrid systems were studied in detail for process feasibility and economic performance. The best-performing system was estimated to produce hydrogen at costs (\$1.67/kg) within Department of Energy targets (\$2.10/kg) for central biomass-derived hydrogen production while also providing value-added energy services to the electric grid.						
15. SUBJECT TERMS hybrid hydrogen; hydrogen system; analysis; biomass pathways; hydrogen production; hybrid; renewable electricity generation						
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